JUNE 2022 | HydrocarbonProcessing.com

HYDROCARBON PROCESSING®

SPECIAL FOCUS: PROCESS OPTIMIZATION

CATALYSTS

PLANT DESIGN, ENGINEERING, CONSTRUCTION AND COMMISSIONING

MAINTENANCE AND RELIABILITY

100TH ANNIVERSARY

CARBON CAPTURE/CO_2 MITIGATION

BIOFUELS, ALTERNATIVE/ RENEWABLE FUELS

HEAT TRANSFER

GAS PROCESSING SUPPLEMENT

DEPARTMENTS

COLUMNS

DIGITAL EXCLUSIVES

WEB EXCLUSIVE

Cover Image: View of BASF's 90,000-tpy acetylene plant at the Ludwigshafen Verbund site in Germany. Around 20 plants at the Ludwigshafen site use acetylene as a chemical building block and starting material for manufacturing many everyday products. Photo courtesy of BASF SE.

HP Industry Perspectives

Hydrocarbon Processing's Mid-Year Update webcast to be held in late June

Over the past several years, the hydrocarbon processing industry (HPI) has been engaged in a whirlwind of peaks and valleys. In the late 2010s, capital expenditures for new refining, petrochemicals and gas processing/LNG capacity were robust, especially in regions such as Asia, Russia, the Middle East and parts of the U.S.

However, the onset of the COVID-19 pandemic and subsequent lockdowns and travel restrictions negatively affected global demand for transportation fuels and several petrochemical product chains. This significant event caused many capital investment projects to be put on hold, reevaluated and even canceled.

As the world emerges from lockdowns, many projects are beginning to restart activities. At the time of this publication, the Gulf Energy Information Global Energy Infrastructure database was tracking more than 1,000 active projects in the refining, petrochemicals and gas processing/LNG industries around the world. A breakdown of active project numbers by region includes the following:

- Africa—102
- Asia-Pacific—437
- Canada—27
- Western Europe—72
- Eastern Europe, Russia and the Commonwealth of Independent States—132
- Middle East—115
- U.S.—146.

These projects will be instrumental in many countries around the world to satisfy increasing demand for certain transportation fuels and petrochemical products, as well as provide natural gas for power generation. In total, these projects represent a total capital investment of more than $1.8 T.

Forecast. What nations are investing in these projects? Why are they investing in new capacity? What major trends are driving the HPI globally? These questions are some of the topics that will be discussed during *Hydrocarbon Processing's* Mid-Year Forecast webcast.

Presented each year, this webcast provides an update to *Hydrocarbon Processing's* Market Data forecast. The webcast, to be held on June 24, will provide global HPI personnel with the latest capital project numbers; major trends, initiatives and company/government programs that are driving change in the industry; and an update on projected spending in capital markets.

More information or to register for the complimentary webcast can be found by visiting www.HydrocarbonProcessing/com/webcasts. HP

HYDROCARBON PROCESSING®
www.HydrocarbonProcessing.com

P. O. Box 2608
Houston, Texas 77252-2608, USA
Phone: +1 (713) 529-4301
Fax: +1 (713) 520-4433
Editors@HydrocarbonProcessing.com

PUBLISHER Catherine Watkins

EDITOR-IN-CHIEF/ ASSOCIATE PUBLISHER Lee Nichols

EDITORIAL

Managing Editor	Mike Rhodes
Digital Editor	Courtney Blackann
Technical Editor	Sumedha Sharma
Reliability/Equipment Editor	Heinz P. Bloch
Associate Editor	Tyler Campbell
Contributing Editor	Alissa Leeton
Contributing Editor	Anthony Sofronas

MAGAZINE PRODUCTION / +1 (713) 525-4633

Vice President, Production	Sheryl Stone
Manager, Advertising Production	Cheryl Willis
Senior Production Manager	Angela Bathe Dietrich
Assistant Manager, Editorial Production	Melissa DeLucca
Graphic Designer	Krista Norman
Technical Illustrator	Joseph Henderson

ADVERTISING SALES
See Sales Offices, page 89.

CIRCULATION / +1 (713) 520-4498 / Circulation@GulfEnergyInfo.com
Director, Circulation — Suzanne McGehee

SUBSCRIPTIONS
Subscription price (includes both print and digital versions): One year $399, two years $679, three years $897. Airmail rate outside North America $175 additional a year. Single copies $35, prepaid.

Hydrocarbon Processing's Full Data Access subscription plan is priced at $1,995. This plan provides full access to all information and data *Hydrocarbon Processing* has to offer. It includes a print or digital version of the magazine, as well as full access to all posted articles (current and archived), process handbooks, the *HPI Market Data* book, Construction Boxscore Database project updates and more.

Because *Hydrocarbon Processing* is edited specifically to be of greatest value to people working in this specialized business, subscriptions are restricted to those engaged in the hydrocarbon processing industry, or service and supply company personnel connected thereto.

Hydrocarbon Processing is indexed by Applied Science & Technology Index, by Chemical Abstracts and by Engineering Index Inc. Microfilm copies available through University Microfilms, International, Ann Arbor, Mich. The full text of *Hydrocarbon Processing* is also available in electronic versions of the Business Periodicals Index.

DISTRIBUTION OF ARTICLES
Published articles are available for distribution in a PDF format or as professionally printed handouts. Contact Mossberg & Co. for a price quote and details about how you can customize with company logo and contact information.

For more information, contact Lisa Payne with Mossberg & Co. at +1 219 561 2036 or lpayne@mossbergco.com.

Hydrocarbon Processing (ISSN 0018-8190) is published monthly by Gulf Energy Information, 2 Greenway Plaza, Suite 1020, Houston, Texas 77046. Periodicals postage paid at Houston, Texas, and at additional mailing office. POSTMASTER: Send address changes to Hydrocarbon Processing, P.O. Box 2608, Houston, Texas 77252.

Gulf Energy

VERIFIED AUDIT PENDING

President/CEO	John Royall
CFO	Ed Caminos
Vice President, Upstream and Midstream	Andy McDowell
Vice President, Finance and Operations	Pamela Harvey
Vice President, Production	Sheryl Stone
Vice President, Downstream	Catherine Watkins

Publication Agreement Number 40034765 Printed in USA

Other Gulf Energy Information titles include: *Gas Processing™, H2Tech, Petroleum Economist©, World Oil®, Pipeline & Gas Journal* and *Underground Construction.*

LEE NICHOLS, EDITOR-IN-CHIEF/ASSOCIATE PUBLISHER
Lee.Nichols@HydrocarbonProcessing.com

Honoring the industry's best: The *HP* Awards

The hydrocarbon processing industry (HPI) is filled with unique and innovative people that stive to provide technologies, processes and services to better the industry in which they work. Their contributions to the global refining, petrochemical and gas processing/LNG sectors are instrumental in providing products demanded by billions of people every day. *Hydrocarbon Processing* believes that these contributions should be celebrated.

Each year, *Hydrocarbon Processing* recognizes the people and companies within the global HPI that are providing new processes, technologies and know-how to advance operations, engineering, construction, safety and sustainability throughout the processing industries. These accomplishments are acknowledged through the *HP* Awards.

_HP_ Awards celebrates ingenuity. *Hydrocarbon Processing* will be celebrating the 5th annual *HP* Awards in-person on October 12 in Houston, Texas. These awards honor the latest technologies and people that have been instrumental in improving facility operations over the past year and making the HPI safer, more efficient and more profitable. These people and companies deserve recognition for their achievements; the *HP* Awards was established to do just that.

Past *HP* Award winners include companies and people from nearly every facet of the HPI. This includes operators, licensors, vendors and suppliers, service companies, and engineering, procurement and construction contractors. These companies/people provide the global processing industries a look into the many benefits and cutting-edge technologies that this industry can produce in a given year. The *HP* Awards provide an outlet to recognize these achievements and celebrate ingenuity.

Categories. The *HP* Awards will feature more than 20 categories, including three categories celebrating industry leaders. *Hydrocarbon Processing* will accept nominations until late June within the categories listed in **TABLE 1.**

For more information about the 2022 *HP* Awards, a complete list of rules/awards methodology, or to nominate your technology or colleague, visit www.HydrocarbonProcessing.com/events for additional details. Finalists for this year's awards will be notified in August. All finalists will be featured within a special section of the October issue of *Hydrocarbon Processing.*

The editors of *Hydrocarbon Processing* look forward to receiving your nominations and wish all submitting companies the best of luck in this year's awards. HP

TABLE 1. ***HP* Awards' categories**

Automation technology	Digitalization technology
Catalyst technology	AR/VR/AI advances
Cybersecurity program/software	Project/asset milestones
Refining technology	Licensor of the year
Petrochemical technology	EPC of the year
Gas processing/LNG technology	Sustainability
Flow control technology	Consulting firm of the year
Health, safety and environment	Lifetime achievement
Instrument technology	Executive of the year
Modeling technology	Most promising engineer

INSIDE THIS ISSUE

K. FINNAN, Yokogawa, Hartford, Connecticut

The acceleration of industrial autonomy and its impact on sustainable manufacturing

In 2020, the author's company published the results of a survey that highlighted a significant shift to industrial autonomy, with as many as 89% of companies planning to increase the level of autonomy in their operations.

Building on the valuable insights of this research, the author's company commissioned a further study in 2021 to gain an in-depth industry perspective on how major manufacturing companies are implementing industrial autonomy. This article summarizes the key findings and insights from that survey.

Survey scope. The survey *Global End-user Survey on the Implementation of Industrial Autonomy* was conducted in September 2021. The survey received 534 respondents from 390 companies across seven global markets and seven different industry sectors. A total of 20 questions examined insights in the following areas:

- The impact of industrial autonomy
- Levels of deployment
- Readiness of technologies
- Decision-making
- The impact on employment
- Challenges.

Respondents were all in management roles within different areas of companies, including corporate; operations, project and plant levels; and information technology (IT). Of the resondents, 62% were manufacturers or process operations, 20% were systems integrators and 18% were original equipment manufacturers. Company sizes ranged from medium to large, with workforces ranging from 1,000 to more than 100,000. A breakdown on industry sectors of the respondents included the following:

- Chemicals and petrochemicals—15%
- Pharmaceuticals and biopharma—20%
- Oil and gas—39%
- Power generation (traditional/fossil fuel)—12%
- Renewable energy—14%.

Key findings. The 2021 survey shed light on several findings that have changed since the 2020 survey. As expected, industrial autonomy deployment has accelerated. The biggest surprise was that environmental sustainability has emerged as an area in which industrial autonomy is expected to make a significant positive impact.

The 2020 survey revealed that 30% of process industry companies had implemented autonomous operations and 30% planned to begin them in the next 3 yr. This trend appears to be accelerating, with 51% of manufacturers surveyed in the 2021 study now scaling deployment of industrial autonomy across multiple facilities and business functions, and another 19% having deployed it in at least one facility or business function.

Further confirmation of this acceleration can be seen in the companies that are not engaging in industrial autonomy. In the 2020 survey, 7% reported that they had no plans to implement autonomous or semi-autonomous operations, whereas all the manufacturing companies surveyed in 2021 are now actively considering industrial autonomy of some kind (FIG. 1).

Aligned with a global focus on the greening of industry and a drive towards more sustainable manufacturing processes and operations, the survey revealed that 45% of respondents expect industrial autonomy to have a significant impact on environmental sustainability improvements in the plant. Only 6% expect industrial autonomy to have no impact on environmental sustainability.

When asked specifically about sustainability, respondents indicated that high levels of positive impact are expected across energy management, worker safety, greenhouse gas reduction and waste reduction.

These findings reveal the developing industry recognition of industrial autonomy as a way to achieve environmental impact reductions and sustainability goals.

Due to the expected positive impacts of industrial autonomy, more than a third (37%) of respondents have already implemented multisite industrial autonomy projects, and 31% have implemented single-site projects, specifically for environmental sustainability, encompassing dynamic energy optimization, water management and emissions reduction.

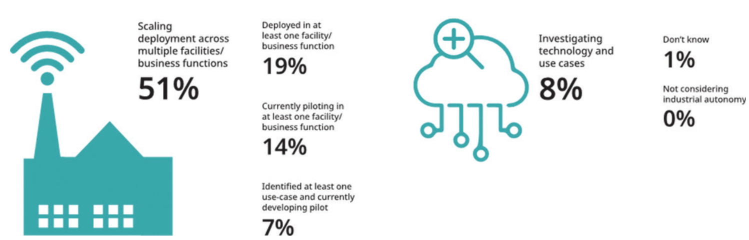


FIG. 1. Companies are starting to deliver on their intentions to deploy industrial autonomy, driving an increase in adoption and implementation, although considerable scope remains for broader deployment.

Return on investment. Productivity improvements in production/manufacturing processes are expected to deliver the highest return of investment (ROI) resulting from investments in digital transformation over the next 3 yr, with 31% of respondents ranking this first and a further 20% ranking it second (**FIG. 2**).

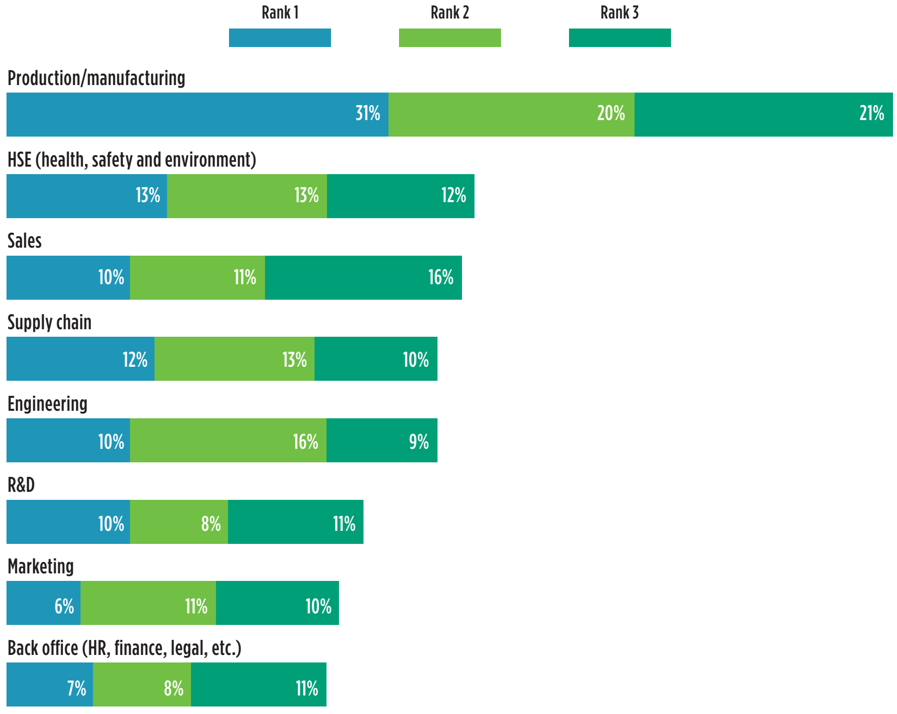


FIG. 2. Improvements in manufacturing and production processes represent the primary area for ROI from investments in digital transformation, but there is a rising significance of ROI related to HSE benefits.

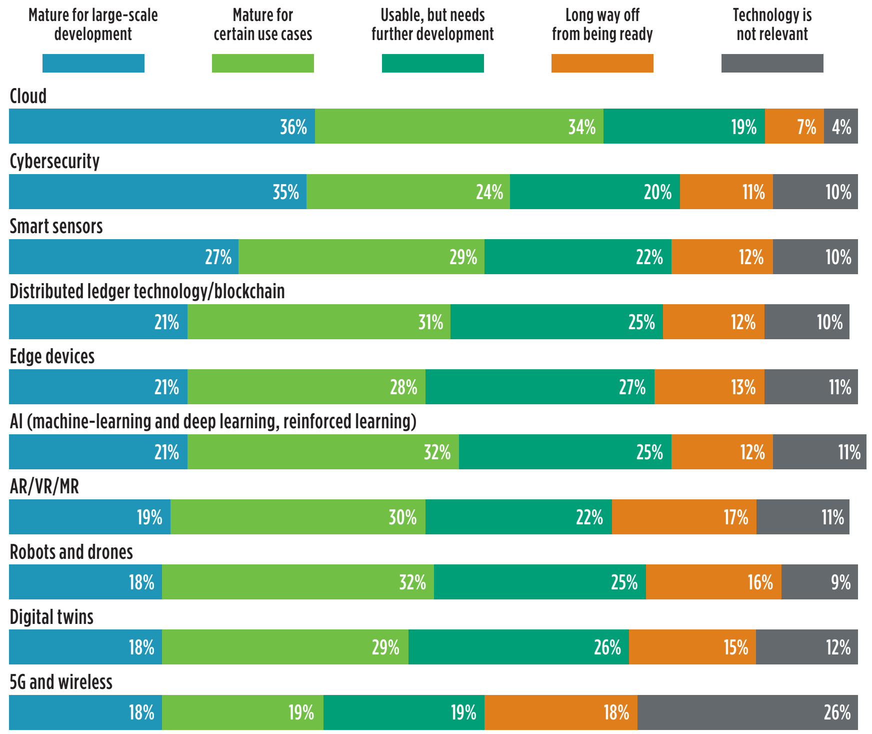


FIG. 3. Technologies are generally ready for industrial autonomy, with cloud leading the way.

However, health, safety and environment (HSE) is emerging as another area of significant ROI, with 26% ranking it first (13%) or second (13%) as the area for greatest ROI resulting from investment in digital transformation in the next 3 yr.

Is the technology ready? Technology readiness is a critical factor in successful industrial autonomy deployments. The survey revealed that the cloud is ranked as the technology area most ready to support industrial autonomy, with 36% of respondents citing that the cloud is mature enough to support large-scale deployment and 34% quoting it as a mature technology ready for certain use cases of industrial autonomy (**FIG. 3**).

Other technologies such as distributed ledger/block chain, edge devices, AI, robotics, and digital twins are most commonly cited as being mature enough for certain use cases.

Remote operations capability is a catalyst, but what is the impact of COVID-19? Increasing the level of remote operations in manufacturing processes represents an important factor in industrial autonomy. The 2020 survey showed that as a direct result of COVID-19, more companies were looking to invest in remote operations. The ongoing impact of the COVID-19 pandemic is likely to continue to drive the adoption of remote working and autonomous operations.

One-third of manufacturers are deploying single-site remote operations and 31% are implementing it across multiple sites in connection to industrial autonomy. Furthermore, 41% of manufacturers expect industrial autonomy to have a significant impact on their remote operations capabilities (**FIG. 4**).

What are the effects on the workforce? The general assumption that increased industrial autonomy, with self-governing systems and higher levels of autonomous operations, will lead to the loss of jobs is challenged by the findings of the survey. According to the respondents, there will be a greater emphasis on industrial autonomy to create new opportunities for the evolution of roles.

Across eight key roles in a manufacturing plant, an average of 40% will re-

quire some retraining, and an average of 28% will require extensive retraining as a result of industrial autonomy implementations. More than 26% will see minimal impact from industrial autonomy or remain unchanged, whereas only 6% of these roles will be eliminated.

The survey also showed that 58% of respondents see industrial autonomy/autonomous operations as systems that augment human tasks, provide decision-making assistance and adapt to changing conditions.

Who are the key decision-makers?

The survey underscored the importance of the C-suite in industrial autonomy decision-making. It showed that the Chief Executive Officer (CEO) plays a key role when it comes to implementing plant level autonomy, with 38% of respondents ranking the CEO as most influential in decision-making. The Chief Technology Officer (34%) and Chief Information Officer (31%) are also seen as key decision-makers, demonstrating the prevalence of C-suite involvement.

Senior-level technical professionals provide important support to these decisions, with 43% saying the Chief Digital Officer has a significant influence on plant-level autonomy decisions.

Industrial autonomy deployments are significant capital investments that fundamentally change working processes. It is clear that decisions are made in the boardroom and are influenced by senior-level technology professionals. HP

KEVIN FINNAN is a Market Intelligence and Strategy Advisor at Yokogawa. He was previously an independent consultant, Vice President of Marketing for CSE-Semaphore and Director of Marketing at Bristol Babcock. He has more than 30 yr of experience in a variety of vertical markets and has launched more than 40 products in automation and measurement technologies.

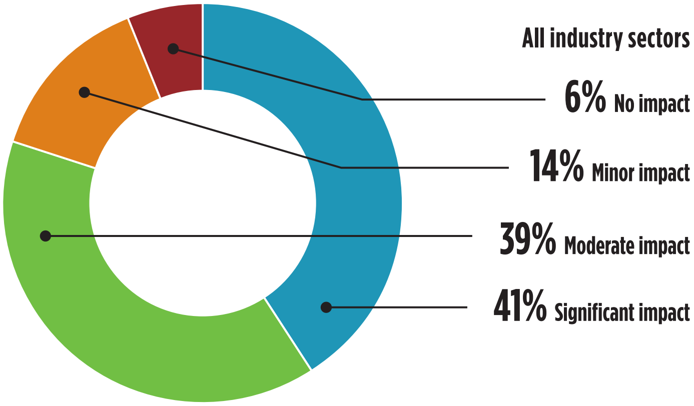


FIG. 4. A greater necessity for remote operations in a post-COVID manufacturing environment is integral to the introduction of industrial autonomy.

MIKE RHODES, MANAGING EDITOR
Mike.Rhodes@HydrocarbonProcessing.com

HP Innovations

Carboline launches CarboNext

In a continued effort to serve the corrosion industry, Carboline has launched CarboNext, a program to educate, connect, support and engage with the next generation of engineers and professionals. CarboNext's core audience includes corrosion professionals 40 years old and younger who are engineers, specifiers, inspectors, project managers, operators or anyone who desires to learn more about coatings and corrosion.

Through outreach and networking opportunities, Carboline strives to be an organization that provides resources to those who look to further their knowledge of corrosion and protective coatings. In addition to training and education, the CarboNext program will include technical events and community support projects.

FIG. 1. The Intelligent Valve from Siemens Smart Infrastructure.

FIG. 2. Sensor Networks Inc.'s 3rd-generation microPIMS® Intrinsically Safe.

Intelligent valve combines energy efficiency, self-optimization

The Intelligent Valve from Siemens Smart Infrastructure (**FIG. 1**), a self-optimizing, dynamic valve with cloud connection, now features an adaptive flow optimizer. This technology, as well as new features such as automatic presetting, continuous optimization of the delta T limitation (ΔT) and system-generated reports, make it possible to achieve cost and energy savings while maintaining comfort for the building occupants. For the first time, actions are autonomously evaluated and executed by the Intelligent Valve and communicated to the building operator.

The company states that the valve is more efficient and competitive than conventional pressure-independent control valves (PICV). Thanks to the automatic preset function, the valve now has the ability to automatically adjust the volumetric flow presetting during operation and to control itself. This offers up to 37% energy savings for cooling and ventilation.

The built-in learning mechanism is based on system demand and avoids maximum volumetric flow when it is not required. This reduces the probability that the delta T (ΔT) limitation control has to intervene, which would lead to a loss of comfort. Due to automatic presetting, installation is quick and easy because the system does not require power and the installer does not have to adjust each valve multiple times. As a result, the valve supports out-of-the-box installation.

The Intelligent Valve's self-adjustment feature makes it easy to accommodate changes of usage and, among other things, helps to continuously optimize the system and immediately detect faults and anomalies. The self-test report makes it easier to document the settings data and prove the efficiency of the system at any time. Since the report is system-generated, entry and copy errors are prevented.

Additionally, the RS458 interface for connection to Modbus RTU networks facilitates integration in facility automation systems and improves monitoring of operating equipment anywhere in the world.

Wireless, non-intrusive ultrasonic corrosion monitoring

Sensor Networks Inc. (SNI) has released its 3rd-generation Permanently Installed Monitoring System, microPIMS® Intrinsically Safe.

This star-network topology system leverages SNI's success and experience in non-invasive corrosion/erosion monitoring with this network of ultrasonic sensors. Intrinsically safe, fully wireless and non-intrusive, it operates using long-range sub-gigahertz LoRaWan® wireless connectivity.

Each microPIMS sensor (**FIG. 2**) can be programmed to take thickness readings at any user-defined time interval. Data is automatically sent to private webPIMS™, cloud-based or on-premise LoRaWAN system + software back-end for analysis, trending and more.

These units can be easily installed temporarily or permanently with either a band clamp or a magnetic base for tanks and vessels. 2 D-sized batteries provide each unit with 15 yr of life at one reading per day. The microPIMS from SNI can also help to lower costs by reducing scaffolding and insulation removal/refitting for internal corrosion monitoring.

The squeaky wheel gets the (biobased) oil

It is often true that "the squeaky wheel gets the grease/oil," both literally and figuratively. In the past, this maintenance activity had to be done with a petroleum-based lubricant. In response to today's emphasis on greater environmental responsibility, Cortec® has provided a biobased oil, the EcoLine® ELP all-purpose lubricant. EcoLine ELP is a high-performance, bio-

degradable soy-based lubricant and penetrant for general purpose use (FIG. 3). It is formulated from natural seed oils and select additives that offer lubricity and performance superior to conventional lubricants. Its excellent performance, biobased nature and low environmental impact make EcoLine ELP a desirable option for environmentally conscious industries or federal agencies and their contractors seeking to comply with mandatory federal purchasing guidelines of the USDA BioPreferred® Program.

EcoLine ELP contains 95% USDA certified biobased content, surpassing minimum biobased content requirements for products falling into the BioPreferred category of multifunctional lubricants. In addition to being biobased, EcoLine ELP is biodegradable, chlorine-free and not toxic to plants—a much more environmentally friendly prospect than petroleum-based products in case of a leak or spill.

EcoLine ELP has shown good performance in laboratory testing compared with several petroleum-based lubricants. When evaluated according to the ASTM D-3233 Falex Pin and Vee Block Test, which demonstrates the extreme pressure properties of lubricants, EcoLine ELP held up under a much heavier load than the other lubricants while avoiding scarring. Compared to WD-40® and CRC 3-36®, it was able to withstand four times the load; compared to LPS 2®, it was able to withstand 10 times the load. Coefficient of friction for EcoLine ELP was also significantly lower than the competition—as low as one-eighth that of LPS 2.

When tested according to ASTM D-4172 Wear Preventive Characteristics of Lubricating Fluid (Four-Ball Method), EcoLine ELP showed a much smaller scar diameter than the three other products. EcoLine ELP has many uses. Chain and gear systems need frequent lubrication and are often out in the open at risk for outdoor leaks and spills where a biodegradable product would be preferred. Nuts and bolts that must be easily removed for equipment access and fans, wheels and hinges are other common moving parts where biobased EcoLine ELP can be introduced.

EcoLine ELP and the three other products were assessed by applying them to rusty nuts and bolts and leaving them to sit for 15 min. They were then unscrewed, and the torque was measured. EcoLine ELP fell into close range with the performance of the competition. Also, although corrosion protection is not the main purpose of EcoLine ELP, parts treated with it passed more than 3,000 hr corrosion-free in 100% humidity at 50°C (122°F) (ASTM D-1748).

Expanded Microsoft Azure machine-learning support

Seeq Corporation, a leader in manufacturing and Industrial Internet of Things (IIoT) advanced analytics software, has announced additional integration support for Microsoft Azure machine-learning. This new Seeq Azure Add-on (FIG. 4) enables process manufacturing organizations to deploy machine-learning models from Azure Machine Learning as add-ons in Seeq Workbench. The result is that machine-learning algorithms and innovations developed by IT departments can be operationalized so frontline OT employees can enhance their decision-making and improve production, sustainability indicators and business outcomes.

Seeq's strategy for enabling machine-learning innovations provides end users with access to algorithms from a variety of sources, including open source, third-party and internal data science teams. With the new Azure machine-learning integration, data science teams can develop models using Azure Machine Learning Studio and then publish them using the Seeq Azure Add-ons feature, available on GitHub. Using Seeq Workbench, frontline employees with domain expertise can easily access these models, validate them by overlaying near real-time operational data with the model results, and provide feedback to the data science team. This enables an iterative set of interactions between IT and OT employees, accelerating time to insight for both groups, while creating the continuous improvement loop necessary to sustain the full lifecycle of machine-learning operations.

Along with increased access to machine-learning models through this integration, Seeq's self-service applications enable frontline employees to perform ad hoc analyses and use the models themselves, rather than rely on an IT team member for support. As the models yield results, Seeq empowers users to scale them across the organization to improve asset reliability, production monitoring, optimization and sustainability.

In addition to launching the Azure integration, Seeq is also expanding its list of published open source algorithms with the addition of a new Seeq Add-on to GitHub for multivariate pattern search. HP

An expanded version of Innovations can be found online at www.HydrocarbonProcessing.com.

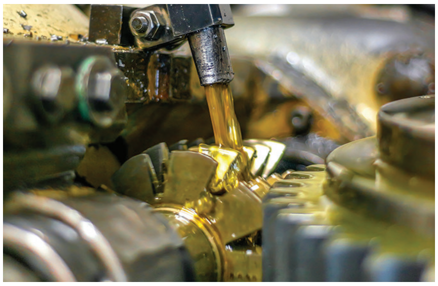


FIG. 3. Cortec® has provided a biobased oil, the EcoLine® ELP all-purpose lubricant.

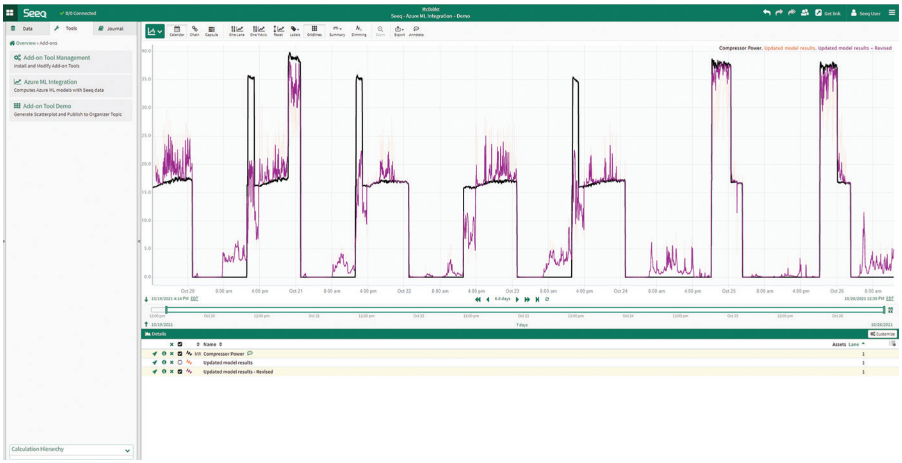


FIG. 4. The new Seeq Azure Add-on enables process manufacturing organizations to deploy machine-learning models from Azure Machine Learning as add-ons in Seeq Workbench.

HEINZ P. BLOCH, RELIABILITY/EQUIPMENT EDITOR
Heinz.Bloch@HydrocarbonProcessing.com

Screw compressor volume control options

During their careers, most rotating machinery experts have dealt with the selection of advanced screw compressors. These are often chosen for their overall simplicity and efficiencies that generally exceed those of centrifugal compressors. Although control options can differ for dry-running vs. liquid injection-type screw compressors, volume control helps control the wide possible operating ranges of these machines.

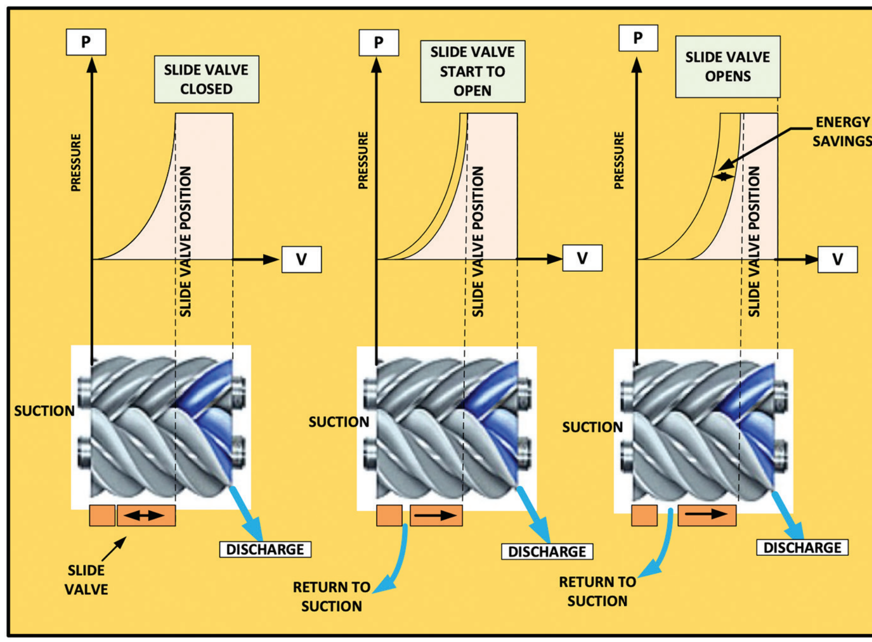


FIG. 1. Slide valve operation.

Controlling dry-screw compressors.

The following options can be used to control dry-screw compressors.[1]

Control by variable speed. Because screw compressors displace the medium positively, the most advantageous volume control strategy is to vary the speed. This may be done by using variable speed electric motors, steam turbine drives, hydraulic or hydro-mechanical torque converters, and other means.

Speed may be reduced to about 50% of the maximum permissible speed. Induced flow volume and power transmitted through the coupling are reduced in roughly the same proportion.

Bypass. Using this method, the surplus gas volume is made to flow back to the intake side by way of a compressor discharge pressure controller. An intermediate cooler brings the surplus gas volume down to intake temperature.

Full-load/idling-speed governor. As soon as a predetermined final pressure is attained, a pressure controller operates a diaphragm valve that opens a bypass between the discharge and suction sides of the compressor. When this occurs, the compressor idles until pressure in the system drops to a predetermined minimum value. The valve will close again on receiving an impulse from a pressure sensor. This brings the compressor back to full load.

Suction throttle control. Suction throttling is suitable for only air compressors. As in the case of the full-load/idling-speed control method, a predetermined maximum pressure in the system (e.g., in a compressed air receiver) causes pressure on the discharge side to be relieved down to atmospheric pressure. Simultaneously, the suction side of the system is throttled down to about 0.15 bar absolute pressure. When pressure in the entire system has dropped to the permissible minimum value, full load is once again restored.

Controlling liquid-injected screw compressors.

The following are pathways to control liquid-injected screw compressors.

Suction throttle control. Since the final compression temperature is governed by the injected oil, a greater range of compression ratios can be accommodated by liquid-injected machines. This permits the main flow volume to be varied within wide limits.

Built-in volume governor. Large compressors are frequently equipped with an internal volume-regulating device. **FIG. 1** illustrates the principle. A slide valve that is shaped to match the contours of the housing is built into the lower part of the housing. It is designed to move in a direction parallel to the rotors, whereby the effective length of the rotors can be shortened. The range of this control mode is typically between about 10% and 100%. Compared with suction throttling, this type of control offers more efficient operation and is often preferred. HP

REFERENCE

[1] Bloch, H. and H. Elliott, *Compressor Technology Advances: Beyond 2020*, DeGruyter, Berlin, Germany, 2021.

HEINZ P. BLOCH resides in Montgomery, Texas. His professional career commenced in 1962 and included long-term assignments as Exxon Chemical's Regional Machinery Specialist for the U.S. He has authored or co-written more than 800 publications, among them 24 comprehensive books on practical machinery management, failure analysis, failure avoidance, compressors, steam turbines, pumps, oil mist lubrication and optimized lubrication for industry. Mr. Bloch holds a BS degree and an MS degree (cum laude) in mechanical engineering from NCE, Newark College of Engineering.

P. LONG
Opportune LLP

Strategic cost management: Uncover supply chain inefficiencies

Controlling costs is critical for any energy company. Firms may spend to grow through acquisition, develop new business, gain market share or invest in new opportunities. However, as a business activity decreases following this period of growth, what happens to the variable costs? If companies have planned strategically, their variable costs fall directly proportional to their activities. Conversely, if they have invested without careful execution of the growth strategy, companies may be stuck with higher variable costs despite the lower activity.

This asymmetrical relationship can be found through comparisons of selling, general and administrative (SG&A) and cost of goods sold (COGS) to a comparison of sales. If the decrease in SG&A and COGS is not proportional to the decrease in sales, then it is worth investigating to uncover the costs in the supply chain. According to the *The CPA Journal,* costs will typically be sticky if adjusting capacity downward in response to a decrease in sales is more challenging. This statement in the context of hydrocarbon supply chain management is further examined here.

Adding capacity: A supply chain dilemma. Firms add capacity during periods of growth for many reasons. According to one Morningstar analyst, "Consolidation makes sense—economies of scale will enable firms to lower supply costs; less fragmentation will ensure more efficient response to price signals and inventory constraints." This line of thinking was held during the early phase of the pandemic by Devon Energy and WPX Energy. They announced a merger in late 2020 to build a dominant position in the Delaware Basin so neither had to slog through the demand uncertainty on their own. A larger economy of scale combined with a rationalization of activity helped to control initial costs.

Companies add capacity to expand their networks to reach new markets or capture greater margins. One company, PBF Energy, looked to broaden its reach of selling finished goods beyond just its West Coast Torrance Refinery. In February 2020, it completed its acquisition of the West Coast Martinez Refinery, which added to the existing capacity in Southern California. The economics can be quite favorable to add value by delivering the finished product to locations closer to customers. By enhancing an existing network, options were now available to optimize cheaper feedstocks from overseas. The acquisition also brought with it additional deepwater marine facilities, product distribution terminals and product storage facilities. The expansion offered more optionality to marketers with increased production, logistics and storage.

Businesses also add capacity because of the specialized logistics that may be required to move a particular product. Crude refiners and specialty chemical manufacturers share this burden. Railroads are not going to provide free railcars that are specialized for hauling the products. So, a high-volume shipper must decide whether to lease or purchase its fleet. Most downstream companies with substantial volume operate fleets of railcars. Though the railroads own some cars, the pendulum has swung to companies owning and leasing fleets of their own (**FIG. 1**). The typical railcar lease is 7 yr; the typical ownership timeline is approximately 30 yr. These are major capital outlays to support a supply chain strategy.

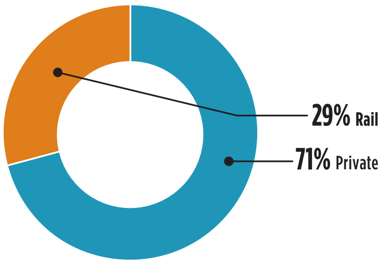


FIG. 1. U.S. freight car by owner. Source: FreightWaves.

Managing the supply chain. In the case of supply chain management, it may appear that expansion and control over your supply chain and logistics are counterintuitive to strategic cost management. As presented in the examples above, that can indeed be the case. In each of the examples, management teams weighed the pros and cons of contracts vs. leasing vs. owning. In the end, control of the inventory and movement of the products won out over cost management—the reason is the overall importance at the time that refiners placed on having complete control over their shipments of cheap crude domestically.

However, this is not to say that you cannot have your cake and eat it, too. It is possible to have strategic cost management if you recognize the vision and strategic imperative behind certain costs. Before you act on expansion, it is important to evaluate the processes that the company will be taking on. For example, when Calumet Superior built a $10-MM rail facility across the street from its refinery, it suddenly entered rail and railcar management.

According to the refinery manager, the movement of crude was "never intended to be a major part of operations." Ask yourself:

- Do you have the core competency and skills needed to perform the necessary functions in both the short- and medium-term?
- Is this a competency that your company wants to develop and grow over time?
- Will it continue to be a competitive advantage, or is the expansion opportunistic and fleeting and will it need to contract

when the market activity reduces?

Such was the case with Philadelphia Energy Solutions, which had to pay $30 MM quarterly to North Yard Logistics—based on the terms of the deal—even if the refinery was receiving little to no crude.

The best protections against long-term costs that burden a company are visibility and transparency. The implementation of a digital strategy allows for the total cost to serve customers to be calculated and evaluated continuously. Digital transformation does not mean having one single enterprise resource planning (ERP) system that replaces all the best-of-breed logistical solutions. Instead, focus on the interfacing of the logistical systems into a central hub where the costs can be analyzed and compared against the sales. Chances are that the commercial deals and transactions may be missing some of the costs to serve customers. Simply put, logistics can make or break the trade.

On the offensive: Strategic cost management. It is important to be strategic in the analysis of the supply chain, rather than purely defensive. In the end, controlling costs is not about gutting your supply chain and disposing of assets at the first sign of a downturn in activity. Rather, it is a continuous insight into the true embedded costs and constant discussion as to how to manage those costs through different commercial terms. It is about developing critical insight into the underlying factors that drive your supply chain and setting up the business processes to maximize the investment. It can also be the foundation for asset-backed trading and increased commercial sophistication.

By maintaining clear insight into the supply chain network and the true cost to serve, management can focus on maximizing profit. The implementation of optimization software is critical to model and create a digital twin of the supply chain. The optimization software ingests the inputs from logistical systems, trading systems and operations, and solves for objective functions like maximizing profit or minimizing costs. Armed with these results, management can now gain insight into their customers and better forecast demand. They can tweak the model to ensure that the supply chain is used most efficiently over time. A strong management team will notice signals and patterns in the data by watching the forecasted demand and analyzing where to take the product for the best netback price. Additionally, by recognizing where the competitive advantage may be lost or market share conceded in a particular geography, a management team can recognize when it must sell down the capacity appropriately. HP

PATRICK LONG is a Director in Opportune LLP's Process & Technology practice. He has more than 20 yr of experience in providing clients with energy trading and risk management, packaged software implementation, trading and risk processes, and business process automation. Mr. Long focuses on studying supply chain issues, implementing digital twins and transforming capabilities to best-in-class solutions. His current focus for clients is making sense of inventory and the supply chain to address management questions, and enjoys thinking of ways to thwart irrational human behavior and its effects on optimization. Before Opportune, he worked in the energy consulting trading and risk systems practice at Accenture, where he managed project teams through the entire process of software selection to the successful implementation of trading and risk management systems for energy trading entities.

Y. ZAK FRIEDMAN
Petrocontrol

Distillation column DCS control configuration

Is there a consensus about a distillation column distributed control system (DCS) control strategy? Much has been written about this topic, covering configuration possibilities, constraints and interacting control considerations. This article proposes a default DCS configuration to serve as a basis. Deviations from the basis are permitted, though reasons for these deviations should be discussed and understood.

Can one DCS configuration be valid for most cases? Distillation columns typically fractionate a mixture of components into a light boiling fraction (distillate) and a heavy boiling one (bottom). The control strategy must first set distillate yield (or cut) correctly to draw all the light components in the feed. This is important because failure to set the cut correctly would contaminate one of the products.

Once the cut is set, the column vapor and liquid traffic (or fractionation) must be set. Separation is not absolute, and cross-product contamination is a function of the internal liquid-to-vapor ratio. It is necessary to define the reflux or reboil ratio to achieve the desired product purities. Therefore, distillation control is about manipulating two handles: cut and fractionation. Precise cut control is more important than fractionation. A 1% cut error may cause significant contamination; whereas, for fractionation, even a 5% shift in the reflux ratio is of minor consequences.

A proposed default configuration that would work for many cases is illustrated in FIG. 1. This DCS structure is called a heat balance-type configuration, augmented by a tray temperature controller (tray TC). The operator sets the reflux flow, whereas the distillate draw is on the accumulator level control. Distillate yield is determined by the heat balance. If reboiler heat duty increases, more vapors will reach the condenser, liquify and be extracted as distillate.

The operator has no precise knowledge of the feed composition and no clue what distillate yield is desirable. Beyond that, feed composition and throughput may vary. Tray TC is a rudimentary inferential controller of the cut. Should the feed become heavier, the tray would warm, and the TC would then cut down reboiler heat duty, reducing distillate yield.

Is a tray TC the perfect inference? No, but it responds in the right direction upon feed composition or enthalpy changes. Without a tray TC, heat balance configurations are difficult to operate. A more elaborate inferential model is feasible, considering more column measurements, internal vapor and liquid traffic. That model could be implemented as a part of the advanced process control (APC) application, and it would reset the tray TC setpoint. However, an elaborate inferential model is not recommended as part of a default DCS strategy.

What about reflux control? Shouldn't the default DCS control strategy include some form of fractionation control? Remember that automatic reflux control is not as critical as cut control because the operator knows what reflux ratio is needed for the separation. Slow ratio controller of reflux to feed or distillate is not out of the question, although dynamic considerations suggest that fractionation control logic should be left for APC. To illustrate the dynamic issues, consider an event when feed flow is ramping up. The tray TC becomes cold, and the controller works hard ramping the reboiler to correct the cut. The last thing needed at that moment is an increase in reflux flow that would further cool the column. The APC would permit increasing the reflux only when the TC is at or above the setpoint.

What about pressure control? The pressure control method should not affect the separation control strategy discussed above, but for completeness, FIG. 1 shows one popular configuration used for total condensers where the condenser is located below the accumulator. This slows down condenser drainage, which builds up liquid levels in the condenser, covering condenser tubes, reducing condenser effectiveness and increasing the column pressure. When column pressure is too

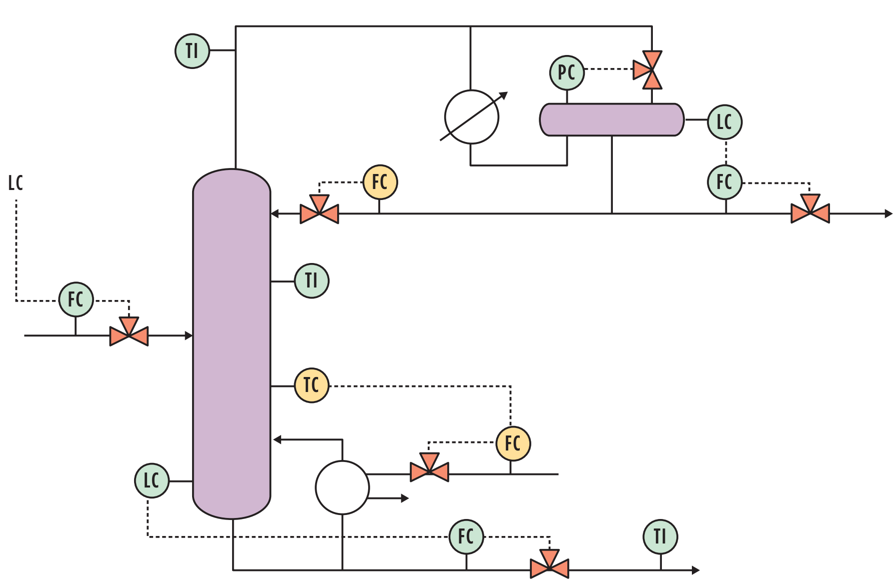


FIG. 1. Heat balance control configuration with stripping section tray TC.

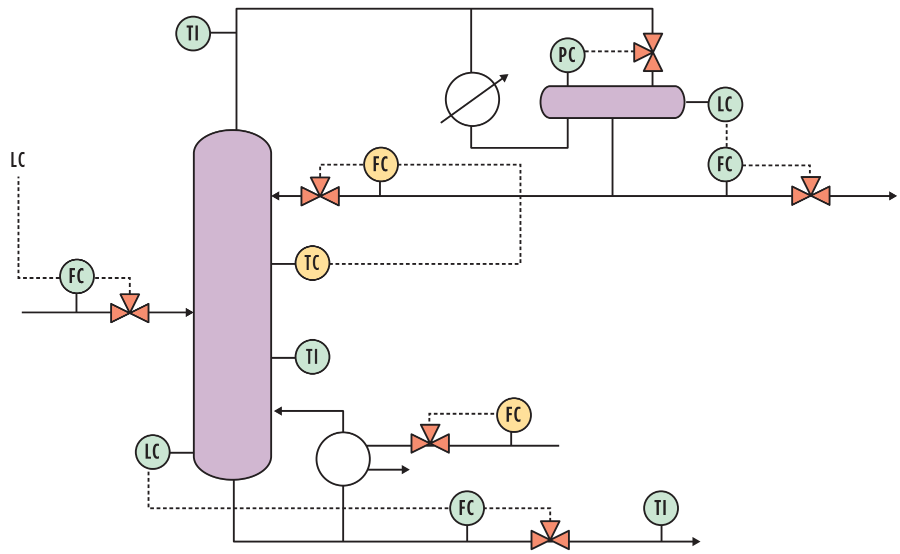


FIG. 2. Tray TC used to manipulate reflux.

high, hot vapor bypass is reduced and accumulator pressure decreases, encouraging condenser drainage, exposing more of the condenser area and increasing condensation to reduce the column pressure. For this concept to work, column pressure must be higher than accumulator pressure, which is accomplished by placing the condenser below the accumulator.

On what tray should the TC be located? **FIG. 1** does not specify whether the TC should be placed near or far from the reboiler. The purpose of distillation is to separate a light component, called a light key (LK), from a heavier component, called a heavy key (HK). Many columns deal with multiple components, some lighter than LK and others heavier than HK. Lighter components would evaporate to become part of the distillate, whereas heavier components would sink to the bottom.

Each column has a composition profile where vapor and liquid composition become heavier near the reboiler. While bottom product specification may call for 1% or fewer LK contamination, the tray TC vapor stream would have a much higher content of LK, ideally 15%–20% for cut control sensitivity. The tray temperature is a function of composition, and upon change, the tray TC would manipulate the reboiler to correct the tray composition. Therefore, the tray TC works by keeping the tray composition constant.

The tray TC location is often determined before control engineers state a preference. Non-ideal locations would hinder the tray its usefulness as a cut control tool.

What about a tray TC located in the rectifying section? Dynamically manipulating the reboiler to control a rectifying section tray temperature is challenging. It is preferred that the TC manipulates reflux flow rather than reboiler duty, as shown in **FIG. 2.** From a heat balance point-of-view, reboiler manipulation is like reflux manipulation. Reducing the reflux would increase distillate yield. Ideally, that TC would be located on a tray where the HK content of the tray liquid would be 15%–20%.

This simple configuration serve as a default configuration for a distillation column DCS strategy. It is clean, simple, non-interactive, easy to operate and keeps the product's purity near-constant. Upon receiving lab results indicating that one product is too clean and the other is too contaminated, the operator would reset the TC setpoint to correct the cut. Conversely, when both products are either too clean or too contaminated, the operator would manipulate the reflux flow setpoint to correct the fractionation. HP

Y. ZAK FRIEDMAN is a Principal Consultant with Petrocontrol. He has practiced APC for most of his career, specializing in using first-principles models for inferential process control and has developed inferential models to fit many hydrocarbon processes. Dr. Friedman's experience spans more than 50 yr, working with Exxon Research and Engineering, KBC Advanced Technology and Petrocontrol. He holds a PhD degree from Purdue University.

A. DORSEY, F. YE and **A. COTTE,** Axens, Rueil-Malmaison, France; and **M. MOLINIER,** Axens, Houston, Texas

Xylene-loop scheme for minimized GHG emissions and PX production cost

Molecular management, energy efficiency and carbon dioxide (CO_2) emissions have become major focuses for petrochemical complexes, including aromatics plants. The xylene loop accounts for about two-thirds of the energy consumption and associated emissions in a modern reformate-to-paraxylene (PX) facility. In the xylene loop, xylene vapor-phase isomerization (VPI) units carry out two chemical reactions: the isomerization of para-depleted xylenes to equilibrium xylenes, and the removal of ethylbenzene (EB) by chemical conversion. EB can either be isomerized to xylenes (the EB-reforming VPI process) or dealkylated to benzene (the EB-dealkylation VPI process). With reformate C_8 aromatic cuts containing up to 18% EB, converting EB to benzene or xylene has a significant impact on the total PX vs. benzene production.

With an unprecedented wave of large steam crackers going onstream in China, Korea, the U.S. Gulf Coast and other parts of the world, pygas benzene is expected to be widely available in these regions. The abundance of benzene supply, combined with potential bans on some benzene derivatives (e.g., bisphenol A in Europe), may drive operators of existing and future aromatics complexes to maximize their PX output while minimizing benzene.

In xylene liquid-phase isomerization (LPI), the only reaction taking place is the isomerization of para-depleted xylenes to equilibrium xylenes. Except in specific situations, LPI is normally deployed in conjunction with VPI to manage EB removal from the xylene loop. LPI operates at a significantly lower temperature than VPI and yields hardly any C_{7-} aromatics and C_{9+} aromatics—therefore, the LPI effluent can be directed to the xylene column without the C_{7-} aromatics fractionation.

A configuration including LPI with an EB-reforming VPI process in an aromatics complex significantly reduces overall product losses, energy consumption and associated greenhouse gas (GHG) emissions, thereby affording the lowest PX production cost. In an existing facility, the addition of an LPI unit also opens the possibility of revamping an EB-dealkylation process into an EB-reforming process. Case studies showing the synergies between LPI and EB-reforming VPI, including the addition of LPI to an existing facility and grassroots facility combining both technologies, are reported here and discussed.

Xylene loop: Two paths for EB removal. Aromatics producers usually define the xylene loop as the circuit that includes the xylene column, the PX separation process, the xylene isomerization process (the effluent of which returns to the xylene column) and the orthoxylene column, where applicable. In facilities converting reformate to PX, the xylene loop accounts for approximately two-thirds of the capital and operating expenses. Consequently, opportunities to reduce PX production costs and utilities consumption primarily rest with the xylene loop design and operation. The C_8 aromatic cuts processed in PX complexes essentially contain three xylene isomers (orthoxylene, metaxylene and PX) and EB. Ironically, PX—the most desired of the xylene isomers—is also the least abundant at equilibrium. Therefore, PX is continuously separated from C_8 aromatic isomers, and, following this separation step, the para-depleted effluent is re-isomerized to equilibrium xylenes and recirculated in a large loop for further PX separation until all xylene isomers are eventually converted to PX. The fourth C_8 aromatic isomer (EB) cannot be fractionated out, since its boiling point is too close to that of xylene isomers—yet it must be eliminated because it takes up space in the xylene loop and reduces the PX production capacity of any facility.

In VPI units where para-depleted xylenes are re-isomerized to equilibrium xylenes, EB is simultaneously removed via a chemical route. Since two options exist for EB removal, two processes are commercially available:

- An EB-reforming VPI process, where ethylbenzene is converted to xylene
- An EB-dealkylation VPI process, where ethylbenzene is converted to benzene.

While aromatics facilities are usually erected to produce PX, benzene is an important byproduct. Naphtha reformate C_8 cuts may contain up to 18% EB.[1,2] EB content in C_8 aromatic streams vary widely depending on the source, with transalkylate C_8 cuts containing as little as 1% EB,[2] and pygas C_8 cuts containing more than 50% EB.[1] In a typical modern plant, reformate represents about half of the xylene-loop C_8 aromatic feed, while transalkylate represents the other half, meaning that the feed EB content is approximately 8%–9%. Therefore, the decision to convert EB to xylene or benzene has a significant impact on the overall feed consumption/PX production capacity of the plant.

FIG. 1A shows the reaction pathway for EB conversion to xylene, and **FIG. 1B** shows the reaction pathway for EB con-

version to benzene. The former reaction is an equilibrium reaction under thermodynamic control—hence, conversion per pass is limited. The latter is an irreversible reaction under kinetic control; in this case, conversion per pass can be adjusted with parameters such as temperature, residence time and hydrogen partial pressure.[2]

State-of-the-art, optimized EB-reforming and EB-dealkylation VPI processes have made substantial progress in minimizing product loss per pass. However, due to the thermodynamically vs. kinetically controlled reaction pathways, the achieved EB conversion per pass is nearly twice as high with EB-dealkylation VPI than with EB-reforming VPI, resulting in higher energy consumption at equivalent EB conversion for the latter process. Conversely, at equivalent PX production, feed consumption will be substantially higher with EB-dealkylation VPI, since EB is converted to benzene instead of xylene. Furthermore, with EB-dealkylation VPI, two out of eight carbons in the EB molecule are converted to light gas, which significantly affects overall plant economics when C_8 aromatic feeds are EB rich.

LPI: Low investment for low product loss and low energy consumption. LPI has been previously discussed in literature.[3,4] This simple and inexpensive process affords lower energy consumption vs. VPI processes—not only because of the absence of feed vaporization and associated hardware, but also because of the lack of side reactions, meaning little to no byproduct generation and a considerably simplified effluent fractionation scheme. Marginal side reactions also mean near undetectable product loss per pass, which is another significant advantage compared to VPI processes. Since the LPI process does not remove EB, it is usually deployed in conjunction with a VPI process to prevent EB accumulation in the xylene loop. The higher the percentage of PX separation raffinate that is processed through LPI vs. VPI, the higher the energy savings and product loss minimization. Conversely, the higher the percentage of PX separation raffinate that is processed through LPI vs. VPI, the lower the overall EB removal. Therefore, the split between LPI-processed and VPI-processed raffinate is the result of a xylene-loop optimization effort.

The two commercial case studies detailed in this article demonstrate that PX production can be maximized via EB-reforming VPI, while both energy consumption and product loss are minimized by the addition of an LPI process. In other words, a scheme comprising EB-reforming VPI and LPI processes effectively leads to the lowest PX production cost and the lowest GHG emissions for a xylene loop.

Case Study 1: LPI process addition to an existing xylene loop operating an EB-reforming VPI process. The flow scheme of this commercial facility is depicted in **FIG. 2.** The plant has opted for EB-reforming VPI to maximize PX production. A continuous catalytic reforming process and a transalkylation process are feeding two xylene loops operated in parallel and comprising a fractionation section including a xylene column and a xylene splitter, an adsorption unit and an EB-reforming VPI unit. The two xylene splitter bottom streams feed an orthoxylene column. The proposed LPI process addition includes two units (shown in green)—namely, one per each xylene loop.

The added LPI heating duty—and marginally increased loop traffic because of the slightly higher EB content—are more than compensated for by the reduction in:

- VPI traffic and associated duty
- VPI effluent fractionation duty
- The xylene column duty.

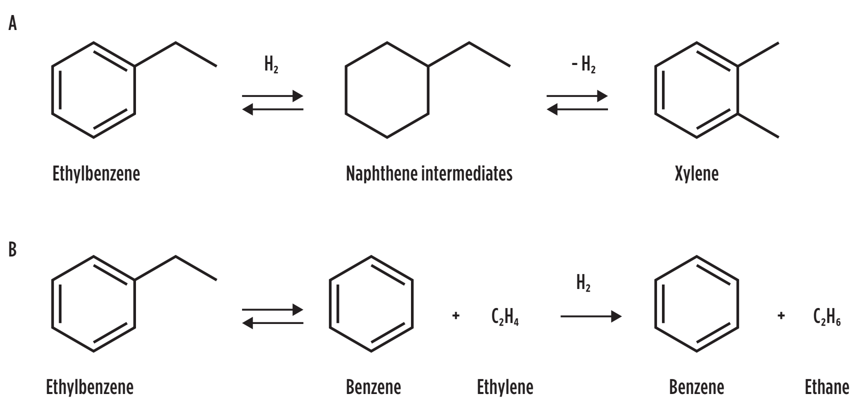


FIG. 1. Reaction pathways for EB conversion to xylene or benzene: (A) EB reforming and (B) EB dealkylation.

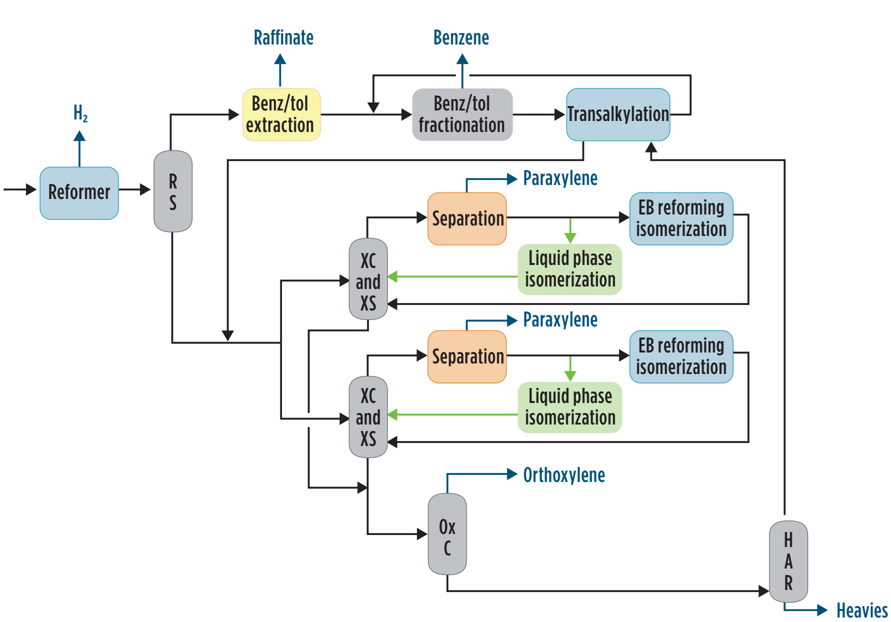


FIG. 2. The addition of two LPI units in an existing facility operating two xylene loops in parallel.

Lower traffic in the VPI process also results in significantly lower xylene losses. Overall energy credits and material balance credits are summarized in **TABLE 1.** In the base case, prior to adding the LPI process, the complex operates with EB-reforming VPI only at design capacity. When LPI is added, the EB-reforming VPI process is set at turndown capacity to maximize the benefits of LPI implementation, which sets the size and feed processing capacity of the LPI unit. At constant PX and orthoxylene production, the xylene-loop feed consumption is cut by ~3%. Another way to substantiate this improvement would be to increase PX production by ~3% at constant feed consumption. Further, and still at constant PX production, electricity consumption decreases by 6% and fuel gas consumption decreases by 2%. The overall PX production cost is reduced by 4% on a $/t basis, and GHG emissions are decreased by 11% on a 1 t CO_2 equivalent/t PX basis. It is worth pointing out that, in an existing facility such as in Case Study 1, the low capital expenditure investment for the LPI addition can be further reduced by revamping onsite equipment, such as converting a clay treater into an LPI reactor or by reusing the previous heat exchanger and/or feed heater, if available. Another important fact is that, at constant PX production, the benefits (such as reduced energy consumption in the reformer and reduced reformate splitter duty) extend well beyond the xylene-loop improvements listed in **TABLE 1.**

TABLE 1. Feed consumption and energy consumption reduction achieved by the addition of two LPI units in an existing facility operating two xylene loops in parallel

Case Study 1	EB-reforming VPI only	EB-reforming VPI + LPI
Xylene loop feed consumption (reformate, transalkylate)	Base	**97%**
PX production	Base	100%
Orthoxylene production	Base	100%
Electricity consumption	Base	**94%**
Fuel gas consumption	Base	**98%**
PX production cost, $/t	Base	**96%**
GHG emissions, 1 t CO_2 equivalent/t PX	Base	**89%**

Case Study 2: Process flow scheme integrating an EB-reforming VPI process and LPI process in the design of a grassroots aromatics complex.

The flow scheme of this planned facility is depicted in **FIG. 3.** One of the main objectives for the proposed plant was to minimize benzene production—therefore, EB-reforming VPI was suggested to maximize PX. Reformate and transalkylate feed a single xylene loop comprising a xylene column, an adsorption unit and

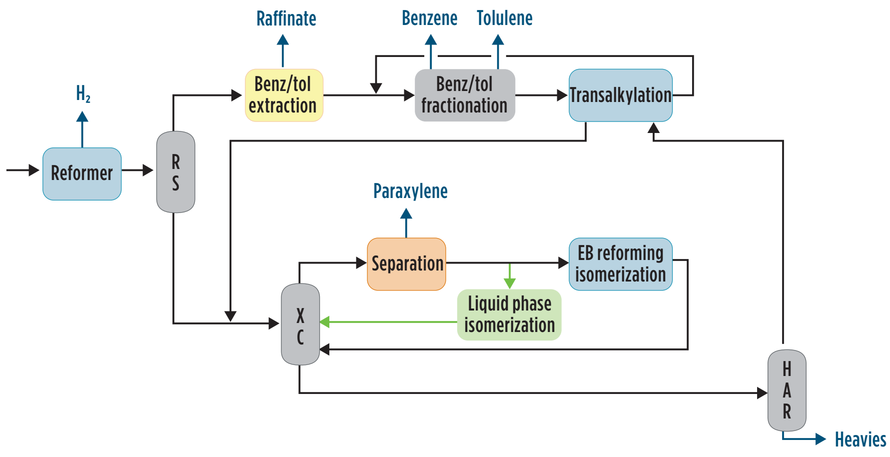


FIG. 3. Integration of an LPI unit in the design of a planned facility with a xylene loop operating an EB-reforming VPI unit.

TABLE 2. Feed and energy consumption reduction achieved by the integration of an LPI unit in a grassroots complex operating an EB-reforming VPI in a xylene loop

Case Study 2	EB-reforming VPI only	EB-reforming VPI + LPI
Xylene loop feed consumption (reformate, transalkylate)	Base	**96%**
PX production	Base	100%
Electricity consumption	Base	**86%**
Fuel gas consumption	Base	**90%**
PX production cost, \$/t	Base	**96%**
GHG emissions, 1 t CO_2 equivalent/t PX	Base	**68%**

an EB-reforming VPI unit. The proposed LPI process addition includes a single unit (shown in green) operating in parallel with the EB-reforming VPI unit.

Overall energy credits and material balance credits are summarized in **TABLE 2.** Energy credits are substantially higher in Case Study 2 (LPI process integration in a grassroots complex) vs. Case Study 1 (LPI process addition to an existing complex). In Case Study 1, the base case heat integration scheme is optimized, while heat integration optimization is limited following the addition of LPI because the constraints associated with existing units cannot be changed.

In Case Study 2, both the base case and the case integrating the LPI process have their equipment size and heat integration scheme optimized—hence, the higher energy credits for the flow scheme including LPI. Feed consumption credits are similar for Case Study 1 and Case Study 2, and they correspond to the expected feed savings associated with the addition of an LPI unit. As pointed out in Case Study 1, feed consumption reduction benefits could also be represented in terms of a production increase at constant feed consumption. However, in an evaluation at constant PX production, feed consumption is cut by 4%, while electricity consumption decreases by 14% and fuel gas consumption decreases by 10%. The overall PX production cost is reduced by 4% on a \$/t basis—an outcome like Case Study 1, despite slightly higher feed savings and significantly higher utilities savings. This is because Case Study 2 was realized in a region of the world where feed and energy costs are substantially lower than in the location of Case Study 1. Conversely, GHG emissions were reduced by more than 30% on a 1 t CO_2 equivalent/t PX basis, owing to a very energy-efficient design, along with substantial reductions in both high-pressure and medium-pressure steam consumption, which are not reported in **TABLE 2.**

Since Case Study 2 was evaluating a grassroots facility, a third case was assessed, combining EB-dealkylation VPI with LPI. As expected, feed consumption at constant PX production increases proportionally to EB content in the fresh C_8 aromatics feed, since EB is converted to benzene instead of to PX in such a scenario. Consequently, of all the possible options, the scheme integrating EB-reforming VPI and LPI processes offers the lowest PX production cost and the lowest GHG emissions/t of PX produced.

Benzene or PX? Benzene and PX prices have been close in recent years,[5] to the point where converting EB to benzene rather than to PX constituted no economic hurdle for most aromatics facilities. This situation could change, however, due to a wave of planned steam cracking facilities—a wave unprecedented by both the number and size of these projects, which includes plants in China, Korea and on the U.S. Gulf Coast.[6,7,8]

Even if pygas benzene production depends on steam cracker feed (i.e., ethane crackers will produce considerably less pygas than naphtha crackers), benzene is still expected to be widely abundant because of the number and size of naphtha cracking facilities. Furthermore, reformers shifting to aromatics mode because of the decreasing gasoline demand are expected to process more benzene precursors, thus delivering more benzene to the market. While benzene becomes increasingly available, the future of some benzene derivatives is presently being reevaluated. For example, a potential ban on polystyrene is under consideration in multiple locations in the U.S.[9,10] Simultaneously, Europe is assessing more restrictive regulations impacting bisphenol A,[11] which may also include future bans. In a world combining increased benzene availability and potentially lower interest in benzene derivatives, converting EB to xylene rather than to benzene—using a combination of EB-reforming VPI and LPI processes—would represent an environmentally friendly and cost-competitive solution for aromatics complexes.

EB-reforming VPI process combined with the LPI process: Additional considerations. The following are additional items for consideration:

- **No PX recovery limitation for existing complexes when adding the LPI process:** The addition of a technology (such as LPI) that is low cost,

and with high energy savings and reduced GHG emissions, is often very attractive to PX producers who are endeavoring to minimize their costs and reduce their environmental footprints. However, there are occasional concerns associated with increased EB traffic in the xylene loop and potentially associated capacity reduction for the PX adsorption process. Experience shows that complexes using the latest molecular sieve technologies have more PX recovery capacity than they can utilize, and that PX capacity limit is a non-issue when adding an LPI unit. The slight increase in EB traffic does not affect the overall PX production capacity, even with the oldest adsorption processes now in operation.

- **LPI effluent circulation around the recovery unit and associated additional savings:** When an LPI unit is added to an existing xylene loop, part of the LPI effluent can be routed directly to the inlet of the PX separation unit (bypassing the xylene column and generating additional energy credits), so long as feed specifications to the separation process are met.
- **EB-dealkylation VPI process retrofit with an EB-reforming VPI process:** Since EB-dealkylation VPI processes operate at a substantially higher weight hourly space velocity (WHSV) than EB-reforming VPI processes, it is unusual to retrofit the former process into an EB-reforming VPI unit. While other process parameters (such as temperature and pressure) would generally be compatible, loading an EB-dealkylation VPI reactor with an EB-reforming VPI catalyst would decrease the overall isomerization capacity of the xylene loop under consideration. This is because there would not be enough room in the EB-dealkylation VPI reactor to fit the required EB-reforming VPI catalyst quantity. However, if such a retrofit is carried out in conjunction with the addition of an LPI process, then the xylene loop can be shifted from converting EB to benzene to converting EB to xylene (and, further, to PX) without incurring feed processing capacity loss and with substantial energy credits gain. Similarly, VPI units that were initially designed to operate an EB-reforming process and that were later debottlenecked via conversion to EB dealkylation, can be converted back to EB reforming at no feed processing capacity loss with the addition of an LPI process to the xylene loop.

Xylene loop for the lowest PX production cost and the lowest GHG emissions. Since feedstock price is by far the most important operating cost parameter in aromatics complex economics,[12] EB-reforming VPI—which converts EB to PX—offers significant advantages to sites predominantly targeting PX product. Adding an LPI process further decreases feed consumption by reducing product losses, while substantially reducing energy consumption and associated GHG emissions. For petrochemical facilities maximizing PX output, a xylene loop combining two specific petrochemical processes[a,b] is the most efficient and environmentally friendly solution for PX production. HP

NOTES

[a] Axens' Oparis® process
[b] ExxonMobil's Liquid Phase Isomerization process

LITERATURE CITED

1 Albahar, M. Z., "Selective toluene disproportionation over ZSM-5 zeolite," doctorial thesis, University of Manchester, UK, 2018, online: https://www.research.manchester.ac.uk/portal/files/75067313/FULL_TEXT.PDF
2 Zhu, F., J. A. Johnson, D. W. Ablin and G. A. Ernst, *Efficient Petrochemical Processes: Technology, Design and Operation,* Wiley, New York City, New York, October 2019.
3 Gonçalves, J. C. and A. E. Rodrigues, "Simulated moving bed reactor for p-xylene production: Dual-bed column," *Chemical Engineering and Processing,* June 2016.
4 Mhatre, S., V. Warke, A. Choi, M. Molinier and A. Saple, "A new liquid-phase isomerization process for xylene loop debottleneck and energy savings," *Hydrocarbon Processing,* January 2020.
5 Statista, "Price of p-xylene worldwide from 2017–2021," online: https://www.statista.com/statistics/1259883/price-p-xylene-globally/
6 Argus, "Viewpoint: Asian ethylene sector braces for new supply," January 5, 2021, online: https://www.argusmedia.com/en/news/2174016-viewpoint-asian-ethylene-sector-braces-for-new-supply
7 Volkova, M., "LG Chem to start up new steam cracker in Yeosu," MRC, June 7, 2021, online: https://www.mrchub.com/news/388673-lg-chem-to-start-up-new-steam-cracker-in-yeosu
8 Stratas Advisors, "North America: Upcoming cracker projects and capacity outlook," April 11, 2019, online: https://stratasadvisors.com/Insights/2019/041119-Downstream-North-America-Upcoming-Cracker-Projects-and-Capacity-Outlook
9 OregonCoast.com, "Polystyrene ban," April 13, 2020, online: https://www.oregoncoast.org/industry/news/polystyrene-ban/
10 Waste360.com, "Washington state bans polystyrene foam, limits single-use plastic at restaurants," May 18, 2021, online: https://www.waste360.com/plastics/washington-state-bans-polystyrene-foam-limits-single-use-plastic-restaurants
11 Mo, C., "Bisphenol A (BPA) regulations in the European Union: An overview," August 25, 2020, online: https://www.compliancegate.com/bisphenol-a-regulations-european-union/
12 Omran, H. R., S. M. El-Marsafy, F. H. Ashour and E. F. Abadir, "Economic evaluation of aromatics production, a case study for financial model application in petrochemical projects," *Egyptian Journal of Petroleum,* 2017.

ANAÏS DORSEY is a Technologist specializing in aromatics and derivatives in Axens' Gasoline and Petrochemicals business line in Rueil-Malmaison, France. She is deeply involved in liquid-liquid extraction and selective reformate hydrogenation applications. She joined Axens in 2014 and has 7 yr of experience in the development of gasoline and aromatics technologies. Ms. Dorsey earned an MS degree in process and chemical engineering from the École Nationale Supérieure des Industries Chimiques (National School of Chemical Industries) of Nancy, France.

FANBING YE is a Technology Manager, specializing in aromatics and derivatives in Axens' Gasoline and Petrochemicals business line in Rueil-Malmaison, France. He is particularly involved in simulated moving-bed PX separation technology. He joined Axens in 2015 and has nearly 7 yr of experience in the development of gasoline and aromatics technologies. Mr. Ye earned an MS degree in general engineering from École Centrale de Lyon (France); an MS degree in automation from Beihang University (China); and an MS degree in refining, gas and engineering from the IFP School of Rueil-Malmaison (France).

ARNAUD COTTE leads the aromatics and derivatives technology group in Axens' Gasoline and Petrochemicals business line in Rueil-Malmaison, France. He has more than 15 yr of experience in the oil and gas industry. Mr. Cotte earned an MS degree in process and chemical engineering from the École Nationale Supérieure des Industries Chimiques of Nancy (France), and an MS degree in refining, gas and engineering from the IFP School of Rueil-Malmaison (France).

MICHEL MOLINIER is a Consultant with Axens North America for aromatics technologies and other petrochemical processes. He has more than 20 yr of experience in the petrochemical industry and 30 yr of experience in heterogeneous catalysis. Dr. Molinier has co-authored two book chapters and 30 peer-reviewed articles and has 25 U.S. patents. He earned an MS degree in physical chemistry from the Université de Bordeaux (France) and a PhD in solid-state chemistry from Philipps-Universität Marburg (Germany).

HP Special Focus

Process Optimization

F.-J. XU, F.-H. ZHOU and **Z.-M. CAO,** China National Air Separation Engineering Co.; **Q. YANG** and **L. ZUBER,** Sulzer Chemtech Ltd., Winterthur, Switzerland; and **J.-J. DONG,** Sulzer Chemtech China, Shanghai

Investigations of high pressure drop observed in ASU columns

The measured pressure drop of columns in air separation units (ASUs) is often reported as being higher than the predicted pressure drop. Particularly for high-pressure (HP) columns, the difference can be up to four times higher, depending on the operating pressure of the column. To investigate the difference, the operation data of recently revamped HP columns were collected; then, vapor and liquid loads as well as their physical properties were generated using a process simulation tool. Thereafter, a proprietary hydraulic rating tool[a] was utilized to predict the dynamic pressure drop caused by the column internals. The use of the differential pressure transmitter and its connecting piping inside the cold box was also scrutinized. It was deduced from calculations that the average temperature of gas in the differential pressure transmitter (PDT) piping located inside the cold box was more than four times higher than the temperature of gas inside the column. The resultant different gas static head at the two locations leads to the deviation between the measured and the predicted pressure drop.

Since the 1980s, structured packings have progressively replaced trays, first beginning with the crude argon (CAR) column and low-pressure (LP) columns. Around 2011, in China, structured packings eventually became a default option for HP columns.

It was determined that the pressure drop of columns in the ASUs was higher than what was predicted by the hydraulic rating tool[a]. The extent of the reported deviations was inconsistent—the typical difference was ~20%, which was initially believed to be within the accurate parameters of pressure drop correlations. When the packings-equipped HP columns began to operate nationwide, the observed deviation jumped. For a 550-kPa HP column, the measured pressure drop was three times higher, and for an HP column operated at 900 kPa, the measured pressure drop was four times higher.

Pressure drop in a column is caused by two factors: dynamic pressure drop and static pressure drop (commonly called gas static head). Dynamic pressure drop is the resistance of column internals, such as packings, to gas flow.[1,2] The pressure drop decreases with declining gas throughput, and reaches zero if gas flow stops. For a packed column, the dynamic pressure drop is closely linked to packings capacity—e.g., for a proprietary structured packing series[b], 5.5 mbar/m and 12 mbar/m correspond to 90% and 100% capacity, respectively. Therefore, the dynamic pressure drop measured from a plant is typically used to monitor packings hydraulic performance. In contrast, the static pressure drop is caused by the pressure exerted by gas weight over a certain height.

The gas density in ASU columns can reach 40 kg/m^3, depending on the operating pressure of the columns, and structured packings generate much less dynamic pressure drop than trays. Therefore, the gas static head can contribute a significant portion to the total pressure drop for ASU packed columns. The proprietary hydraulic rating tool only calculates dynamic pressure drop, so the gas static head must be manually added to the dynamic one to obtain the total pressure drop of a column. However, the measured pressure drop in ASU plants matches neither the proprietary hydraulic rating tool's[a] predicted dynamic pressure drop nor the total manually calculated pressure drop. Rather, it lies between.

Although the deviation does not cause any practical challenges to column design and operation, a joint investigation between the authors' companies began in early 2020 with the objective of benefitting the ASU industry.

Data collection and analysis. Among the various columns in an ASU, the HP column is the best option for investigation: the simplest HP column consists of only one feed (e.g., the compressed air feed) and two products (i.e., liquid nitrogen at the top and oxygen-enriched liquid at the bottom). The high pressure drop observed in HP columns will not be masked by the inaccuracy of any pressure drop correlations.

In the last 2 yr, the authors' companies have successfully revamped HP columns from tray to structured packings, and the details of the first project were published elsewhere.[3]

The operation data of two identical revamped HP columns in a plant were collected in June 2021. The vapor and liquid loads and their physical properties were generated using a process simulation tool, and then the dynamic pressure drop of packings was calculated using the proprietary hydraulic rating tool[a]. The gas static head inside the HP column was calculated manually using Eq. 1:

$$\Delta P_{static\ pressure\ drop\ in\ HP} = \rho_g \times g \times h \quad (1)$$

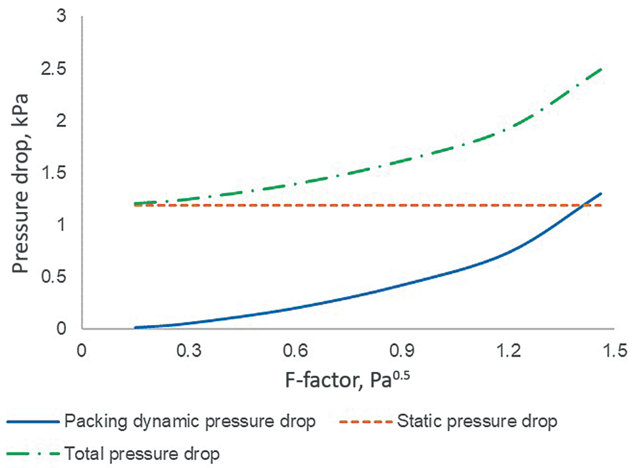


FIG. 1. Pressure drop of the top bed against F-factor.

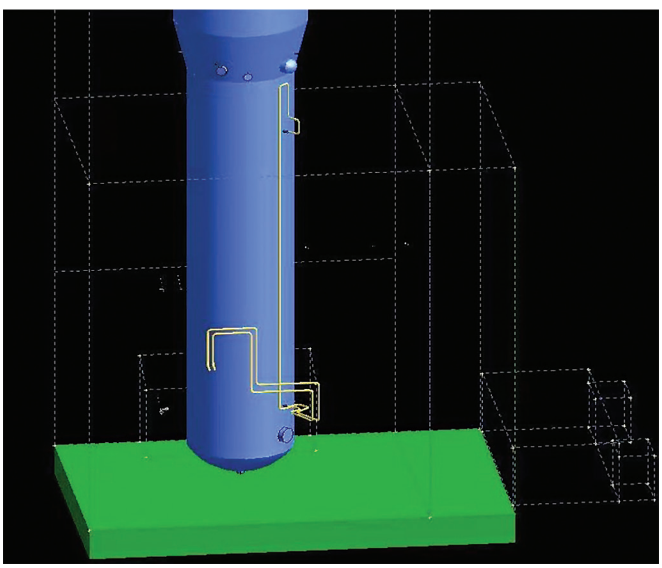


FIG. 2. PDT taps and piping connected to the HP column in the cold box (other equipment is omitted for clarity). Note: the dotted lines indicate the contour of the cold box.

where,

ρ_g is the average gas density in the HP column,
g is the gravitational acceleration and h refers to the total vertical distance between the pressure taps.

The relevant data are listed in **TABLE 1.** The measured pressured drop is about three times the dynamic pressure drop predicted by the hydraulic rating tool for both units. It can also be seen that the gas static head contributes to > 70% of the total pressure drop in the column, as the gas density is between 22.7 kg/m^3 and 22.4 kg/m^3 in the 14-m tall columns.

The HP columns investigated consist of two beds, and **FIG. 1** illustrates three different types of pressure drop for the top bed against F-factor, which can be calculated according to Eq. 2:

$$\text{F-factor} = (V/A) \times \sqrt{\rho_g} \qquad (2)$$

where,

TABLE 1. Pressure drop for the HP columns at the same load

	Unit 1	Unit 2
Predicted dynamic pressure drop, kPa	1.111	1.111
Predicted gas static head in HP, kPa	2.852	2.852
Total pressure drop, kPa	3.963	3.963
Measured pressure drop by PDT, kPa	2.96	3.16

V stands for volumetric flowrate of gas (m^3/sec),
A for the tower cross-sectional area (m^2) and
ρ_g for gas density (kg/m^3).

The change in gas density is < 1% when the operating pressure of the HP column varies between 545 kPa and 551 kPa (the result of change in throughput). Therefore, it is reasonable to assume that the gas density and the subsequent static pressure drop do not vary with the F-factor, as shown in **FIG. 1.**

Using two pressure transmitters, the pressure drop in a column can be measured by subtracting the top pressure from the bottom one. The advantage of this arrangement is its simplicity, as the two transmitters (top and bottom) do not need to be connected by a vertical line. However, a disadvantage can be inaccuracy. In this instance, the typical error of 0.1% on the pressure measurement for a column operating at 550 kPa is expected to be +/– 0.55 kPa for both the top and the bottom measurements. Therefore, the pressure drop difference will have an error of +/– 1.1 kPa. A measurement of 3.0 kPa (as shown in **TABLE 1**) with an error of +/– 1.1 kPa is useless. This method is inadequate, and not recommended in this case.[4] A differential pressure method is preferred.

A proprietary transmitter model[c] was used in the plant. It is known that, in general, a PDT reading does not indicate the gas static head in a column, as the gas static head in the PDT connecting piping would offset the one in the column. However, in some situations, corrections must be made when the PDT reading has to be rigorously interpreted. For instance, in many applications, inert gas such as nitrogen (N_2) is used to purge the piping to prevent condensation of process gas due to lower temperature outside the column. The detailed correction method can be found in literature.[1]

Since the pressure drop measured by the PDT in the ASU was much higher than the predicted dynamic one, corrections were conducted to verify whether the deviation can be explained. A particularity of cryogenic distillation (as found in an ASU) is that the temperature of the piping connecting to the PDT is higher than the temperature inside columns, thus no inert gas needs to be injected into the piping. Instead, process gas enters the pressure piping directly.

To evaluate the impact of the gas static head on the PDT reading, the space arrangement of PDT piping connected to the HP column in the cold box was retrieved from installation drawings. The piping in Unit 1 is shown in **FIG. 2** and simplified in **FIG. 3.** The representative piping with numerous bends counters thermal expansion and contraction in a range of more than 200°C (392°F) between normal operations and startups/shutdowns. The transmitter is on the lower platform for ease of access ($h1$ at 10,465 mm and $h2$ at 2,400 mm). On the day when the operation data were recorded, the ambient temperature was 27°C (81°F). It should be noted that that the

TABLE 2. Static gas head at different temperature of PDT piping

Assumed T in PDT connecting piping, °C	-80	-60	-40	-20	0	20	30
Gas density of N_2 in the piping at 555 kPa, kg/m^3	9.81	8.85	8.07	7.41	6.86	6.38	6.17
Static head in PDT connecting piping, kPa	1.24	1.12	1.02	0.94	0.87	0.81	0.78
Predicted PDT value, kPa	2.72	2.85	2.95	3.03	3.1	3.16	3.18

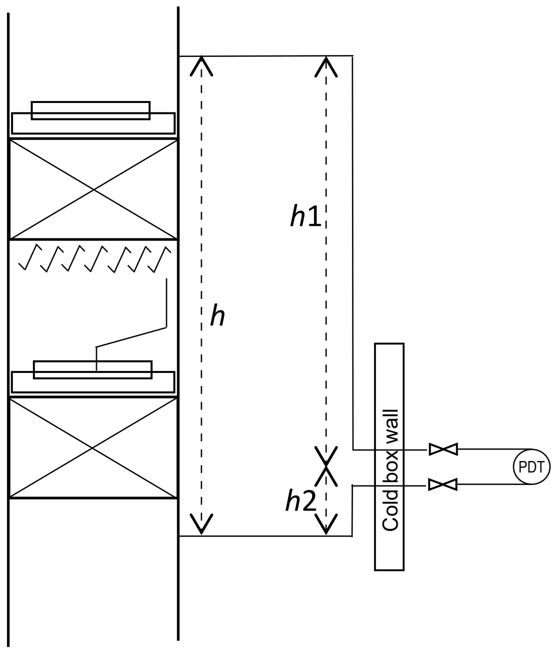


FIG. 3. The simplified HP column and PDT piping.

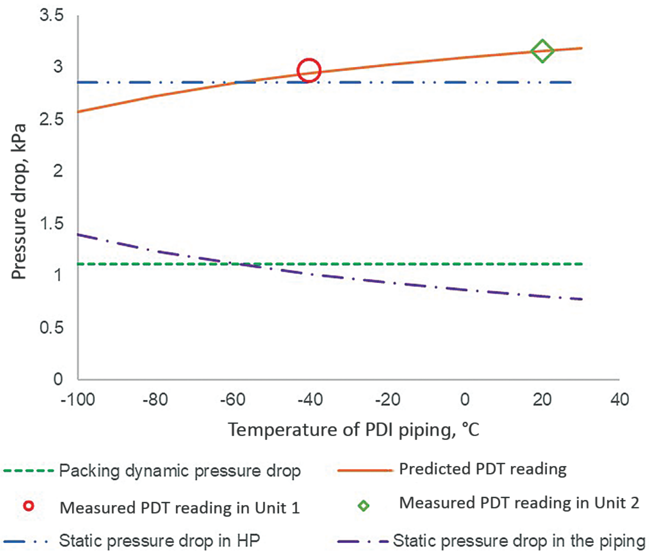


FIG. 4. Pressure drop of the HP column against the temperature of PDT piping at a capacity of 20 kNm^3/hr.

detailed pressure piping design may vary among licensors.

Unfortunately, like in many other plants, the PDT piping local temperature was not available despite the fact that the temperature of the cold box foundation is closely monitored. A different approach was taken in corrections: match back the temperature of the piping based on the measured pressure drop.

A large temperature gradient between the column and the wall of the cold box was expected, as the temperature inside the HP column was close to –180°C (–292°F), in contrast with the 27°C (81°F) ambient temperature. By varying the temperature, the gas density of a pure N_2 at 5.51 bar was calculated and is tabulated in **TABLE 2.** As shown, the gas density changes with temperature, and its values in the temperature range investigated are at least two times smaller than the gas density in the HP column. Similarly, the density of the O_2-enriched gas is also calculated at the corresponding temperature. The static pressure drop in the piping is calculated based on Eq. 3:

$$\Delta P_{static\ pressure\ drop\ in\ the\ tubing} = \rho_{g1} \times g \times h1 + \rho_{g2} \times g \times h2 \quad (3)$$

where,

ρ_{g1} and ρ_{g2} refer to gas density in the upper and lower piping, respectively. Then, the predicted pressure drop by the PDT is obtained by correcting the dynamic pressure drop (Eq. 4):

$$PDT = \Delta P_{packing\ dynamic\ pressure\ drop} + \Delta P_{static\ pressure\ drop\ in\ HP\ column} - \Delta P_{static\ pressure\ drop\ in\ the\ piping} \quad (4)$$

When the piping temperature is assumed as –40°C (–40°F) and 20°C (68°F) for Units 1 and 2, respectively, the calculated PDT values match the site PDT readings at a unit capacity of 20 kNm^3/hr.

It is worth mentioning that in the two identical units, the main equipment and their space arrangements are exactly the same, but the space arrangements of instrument piping are not.

FIG. 4 exemplifies the impact of the temperature of PDT piping on different types of pressure drop. It can be seen that the predicted PDT reading rises with increasing temperature in the PDT piping, and the slope is steep at lower temperatures. It is also apparent that the piping for Unit 2 is closer to the wall of the cold box. The calculations at other loads deduced the same temperature for each unit, confirming the plant data acquired are consistent and of quality.

Assuming a uniform piping temperature in the calculations above is correct for qualitative analysis. However, considering the isometric of the pressure piping in the cold box and its variations in different projects, precise quantification of the gas static head in the piping requires temperature gradient along the piping.

In addition to the variation associated with the temperature gradient along the piping, the location of the transmitter outside the cold box—which is also project dependent—affects the gas static head in the piping to a small extent. Specifically, in the case investigated here, the gas in the upper piping is nearly pure N_2 with trace of oxygen (O_2) and argon (Ar), and the gas in the lower piping consists of 37.8% O_2 and 1.6% Ar. As the gas density is a function of its composition, if the location

of the transmitter changes, then the gas static head in the piping would also change, according to Eq. 3.

Due to a very different temperature in its connecting piping, the measured pressure drop by a PDT in an ASU cannot be used to judge packing hydraulics permanence directly.

Inferences. Despite being a theoretical exercise, the calculations in this paper are adequate to conclude that the deviation between the predicted dynamic pressure drop and the PDT reading is caused by the difference between the gas density in the HP column and the gas density in the pressure piping due to the great temperature difference. The larger the difference in gas density, the greater the deviation.

This can also explain why only a 10%–30% deviation has been observed with the LP and CAR columns. First, the gas density difference is still the primary reason for the deviation. Due to much lower operating pressure, the gas density in the LP and CAR columns are in the range of 5 kg/m^3–8 kg/m^3; therefore, the gas static head becomes a smaller portion of the total pressure drop (the curves of static and total pressure drop would shift downwards in **FIG. 1** in such a case). Using gaseous N_2, **FIG. 5** illustrates the impact of column operating pressure on the gas density at different temperatures.

Takeaway. Due to the temperature gradient along the piping of the differential pressure transmitter and its varying position in the cold box, it would be challenging—if not impossible—to deduce from a PDT reading either the dynamic pressure drop of packings or the total pressure drop of a column. Hence, unlike in other applications, PDT readings in an ASU cannot be used to judge how far the columns are from flooding, nor can they provide any input to process simulation and upstream equipment sizing, significantly limiting typical uses. To obtain the correct pressure drop for ASU columns, zero calibration[5] can be considered based on gas density. However, this requires the gas temperature in the pressure piping.

To investigate the difference between the PDT reading and the predicted pressure drop, the operation data of HP columns in the ASU were collected and processed. It was determined the deviation is caused by significant temperature difference between the HP column and the differential pressure transmitter piping. The PDT reading corresponds neither to the dynamic pressure drop of packings nor the total pressure drop of the column. HP

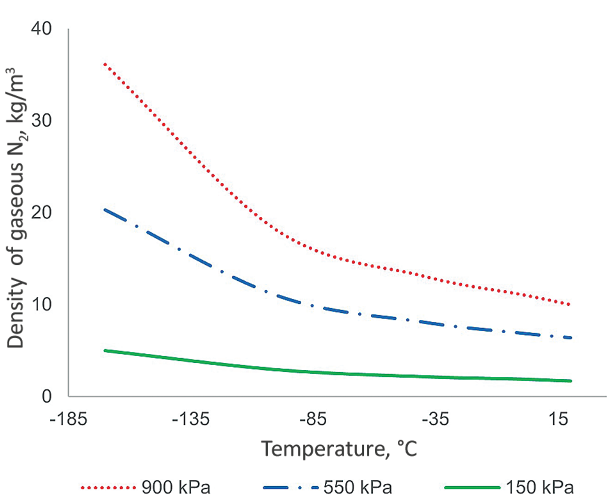


FIG. 5. Density of gaseous N_2 at different temperature and pressure.

NOTES

[a] Sulzer's Sulcol™
[b] Sulzer MellapakPlus™ series
[c] YOKOGAWA EJA110E

LITERATURE CITED

1. Cai , T. J. and M. R. Resetarits, "Pressure drop measurements on distillation columns," *Chinese Journal of Chemical Engineering*, 2011.
2. Kehrer, F., L. Spiegel, E. Kolesnikov and P. Choo, "Experimental investigation and modelling of Sulzer I-ring hydraulics," *Chemical Engineering Research and Design*, Vol. 84, Iss. 11, November 2006.
3. Yang, Q., J. J. Dong, F. J. Feng and X. D. Zhang, "Revamping an air separation unit," *PTQ* , 2019.
4. Kister, H. Z., "Distillation operation," McGraw-Hill Inc., New York, 1990.
5. American Institute of Chemical Engineers (AIChE), *Tray distillation columns: A guide to performance evaluation,* 2nd Ed., Wiley, New York, New York, 1987.


FENGJIE XU is a Senior Engineer at China National Air Separation Engineering Co. Ltd., where he is responsible for R&D of ASU processes and ASU energy-saving technologies. He has been engaged in the field of air separation for 15 yr, and holds a BE degree in energy and environmental engineering.


FENGHUA ZHOU is a Senior Engineer at China National Air Separation Engineering Co. Ltd. She specializes in the automation design and R&D of control technology for ASUs. She has more than 19 yr of experience in the field of air separation automatic control and holds an MA degree in control engineering.


ZHIMING CAO is a Senior Project Manager at China National Air Separation Engineering Co. Ltd. with responsibility over equipment integration, construction, erection and project management. His 21 yr of process project management experience ranges from engineering design, equipment installation to commissioning, etc., for ASU projects.


QUAN YANG is a Sulzer Senior Technical Expert, specializing in mass transfer components. He is in charge of air separation knowledge management within Sulzer Chemtech. Recently, his field expanded to process optimization for DWC. Dr. Yang holds a PhD in environmental engineering from the National University of Singapore.


LAURENT ZUBER heads Process Innovation and Technology Management at Sulzer Chemtech Ltd. in Winterthur, Switzerland. He joined Sulzer in 1995 as a Process Engineer for chemical and petrochemical separation units, and has held different positions in application management. He has a special interest in advanced distillation solutions and difficult product separation. Dr. Zuber holds a PhD in chemical engineering from the Swiss Institute of Technology of Zurich.


JIAO-JIAO DONG is a Principle APT Engineer at Sulzer Chemtech China in Shanghai, China. In her 13 yr with Sulzer, her main responsibilities have included driving ASU business in the China market and advocating the best designs and practices of column internals optimized for ASUs. She holds an MS degree in chemical engineering from the East China University of Science and Technology.

J. HANNON, ExxonMobil Fuels and Lubricants, Allentown, New Jersey

Advanced cleaner technology: Leveraging tailored solutions for reliability goals

Operators at refineries, petrochemical plants and power generation companies are under constant pressure to reduce maintenance and operational costs, boost productivity, minimize risk of unplanned outages and maximize time between overhaul periods. As a result, many oils used in commercial settings are engineered to provide extended performance in severe conditions, with the goal of keeping critical assets online. However, non-oil related events (e.g., localized hot spots) can still lead to significant problems associated with oil oxidation, which impedes overall oil performance, protection and reliability.

In response, advanced oil-soluble cleaner technology has emerged as a promising solution to address these issues and to provide operators with peace of mind when it comes to reliability. However, taking advantage of this technology is not as simple as applying an oil-soluble cleaner product every time there seems to be an equipment issue. Operators should be informed about the full potential of the technology itself, and how it should be leveraged for their unique operational goals.

Leveraging advanced cleaner technology to improve reliability and cut costs. Advanced oil-soluble cleaner technology can help address many issues that hinder equipment reliability, but it is important for operators to understand the differentiation between the solutions available to them.

For example, some of the products on the market for issues like high bearing temperatures and deposit buildup only offer temporary fixes, with the potential for additional problems down the line, including anything from risk of severe deposit buildup and phase separation to accelerated degradation and potential incompatibilities that impact foam and demulsibility. While most oil-soluble cleaners can help with conversions between oil changes and offer some aid in troubleshooting unplanned threats to equipment reliability, few are able to accomplish both effectively, while mitigating the risk of future issues and even extending the life of the current system oil.

When it comes to identifying a solution that is both seamless and highly effective for long-term operation, this is where a product's technology becomes effective. For example, the author's company's proprietary oil-soluble cleaner[a] stands out from other oil-soluble cleaners on the market based on its technology-enabled benefits. The oil-soluble cleaner[a] is based on proprietary technology[b], which enables it to be a high-performing solution designed for power generation and petrochemical equipment. For example, the product is engineered to be compatible with most API Group 1–4 lubricants and hydraulic oils, so it has no adverse impact on in-service oils. Therefore, it can used as a leave-in cleaning solution during oil changes and systems facing issues like high bearing temperatures and varnish buildup without requiring those systems to be taken offline. The proprietary oil-soluble cleaner can also help with oil life extension, allowing operators to have the protection they need to mitigate future issues with performance. Additionally, it can be used in these scenarios at a much lower concentration that most competitors, with only 3%–5% of total oil volume typically added to these systems.

Advanced oil-soluble cleaners continue to show proven success in helping companies across various industries avoid operational shutdowns, improve oil performance, and prevent future and reoccurring issues. However, it is important that operators pay close attention to long-term performance to avoid causing problems beyond the immediate issue. Technical experts can assist in implementing an effective condition monitoring program to monitor long-term performance.

The proprietary oil-soluble cleaner has led to cost savings, as well as helped avoid potential equipment shutdowns by assisting in conversions, troubleshooting ongoing issues and extending current oil life. For example, a refinery was replacing oils and wanted a clean system for the new charge. The author's company's engineering team and reliable industrial group recommended a varnish flush with the proprietary oil-soluble cleaner[a] in lieu of a traditional flushing method, which led to savings of $104,000 on two turbine fills. In another instance, a company in the petrochemicals industry had high-speed cryogenic compressors with frequent bearing temperature spikes that were stabilized within 30 min of the technology being added to the system. Another plant was able to extend compressor oil service life to 3 yr with the use of the proprietary oil-soluble cleaner to mitigate prior varnish issues.

Preventing downtime in critical moments and planning. Sometimes, operators must quickly troubleshoot an issue to prevent critical equipment downtime and avoid a major shutdown that could cost the company millions of dollars. While a rapid fix in these moments is crucial to prevent high costs, it is also important for operators to plan and mitigate the need for aggressive, short-term solutions down the line—resulting in more savings for the operation over time.

For example, the product solutions used should also extend equipment performance and reliability. In addition, operators

should rely on technical expertise from their lubricants provider to implement the right products and maintenance procedures to get ahead of the next critical situation, cutting operational costs in the long term.

In 2021, a refinery requested the author's company's assistance in troubleshooting progressive issues with rising bearing and oil temperatures, as well as water contamination on a critical air blower after an unexpected oil cooler failure. At the time the problem was reported, temperatures had risen to 227°F (108.33°C) and were rising approximately 2°F/d. If these bearing temperatures were not stabilized, the refinery would face a potential unplanned shutdown (30 d–60 d outage), which would lead to a significant cost and loss of productivity.

Through a bearing inspection, the author's company's engineering team was able to identify the problem causing these temperatures—contaminated oil and varnish buildup. Through fast action and the team's recommendations to implement the proprietary oil-soluble cleaner-based technology for its varnish dissolving properties, the refinery was able to identify the root cause of bearing temperature escalation and oil contamination, stabilize temperatures to avoid costly refinery shutdown and maximize productivity. By avoiding this shutdown, the refinery was able to enhance safety, while avoiding costly workforce and equipment costs, saving the company nearly $8.3 MM.

The proprietary oil-soluble cleaner can not only be added to systems suffering from reliability problems, but can also be deployed proactively to minimize varnish deposits and increase reliability. Its unique chemistry provides a long-term solution to deposit formation, even in thermally stressful applications.

Marrying tailored solutions with technical expertise. As the industry continues to evolve with new technologies to maximize efficiency, operators must find tailored solutions that will address issues specific to their operation rather than incorporating a one-size-fits-all solution. No matter how effective, the latest technology innovation is not the answer to all operational problems. Partnering with technical experts that will advise on the right solution for an operator's unique goals is equally important as the most advanced product solution.

For example, the author's company's engineering team and its integrated services partners make personalized recommendations for companies on what actions should be taken to get them the best results in the most efficient way.

Partnering with technical experts can also help operators find a solution tailored to their needs, rather than adopt a quick blanket fix. For example, while oil-soluble cleaners can be somewhat effective for common issues, there are many scenarios where an issue needs a more in-depth cleaning solution, with the bonus of extended oil service life. Therefore, implementing advanced cleaner technology must be coupled with the right technical expertise so operators can be sure they are utilizing the best solution for the problem at hand.

Operators should look for these custom solution plans from their lubricants provider, weighing all options and reviewing potential outcomes. For example, the proprietary oil-soluble cleaner[a]—like any product or service—is not a one-size-fits-all solution but rather a tailored solution for the problem a company is facing. These experts should work diligently to properly diagnose these issues, finding the best solution for the immediate issue at hand, while also aiming for long-term operational success. It is also important for a lubricants provider to leverage partnerships with system cleaning industry leaders to allow for a seamless integration when offering solutions.

A technical team should also use various tools to formulate the proper diagnosis with solution options. For example, the author's company created a turbine triage process to help show customers various maintenance scenarios and where their current issues fall. While some companies may take only one or two oil analysis tests, the author's company's engineering teams look at oil service hours, equipment application suffering points, visual inspection and a comprehensive lube analysis test slate to gain the full picture.

Ultimately, no technology or product can help reduce maintenance, boost productivity or minimize risk. The key to finding the right solution is to have the technical expertise behind it, with a team that works side-by-side with operators to define both short-term and long-term goals and find the best ways to achieve them. HP

NOTES

[a] Mobil™ Solvancer®
[b] Fluitec® DECON™ technology

JIM HANNON is Mobil Industrial's turbine oil Technical Lead. He is a graduate of the Massachusetts Maritime Academy and has worked in the power and marine industries for more than 30 yr, with more than two decades at ExxonMobil.

N. PONGBOOT, Global R&D, Samut Prakan, Thailand; and **T. UPIENPONG** and **W. KARUNKEYOON,** PTT Global Chemical, Rayong, Thailand

Hydroprocessing catalyst selection—Part 2: Considerations and best practices for catalyst testing

Hydroprocessing catalysts are an essential part of the refinery involved in the treatment/conversion of most petroleum fractions, ranging from naphtha to residue. Hydrotreating catalysts help refiners meet fuel regulations and enhance the performance of downstream catalysts and processes, [e.g., naphtha reforming or fluid catalytic cracking (FCC)] by removing sulfur (S), nitrogen (N) and metals from their feedstocks, as well as improving product properties by hydrogen (H) addition. Moreover, hydrocracking catalysts further improve refiners' profits by converting low-value streams [e.g., vacuum gasoil (VGO)] into high-value fuels and chemical feedstocks. Therefore, the selection of hydroprocessing catalysts requires great care from refiners to ensure maximum asset utilization and profitability.

Part 1 of this article, which appeared in the May issue of *Hydrocarbon Processing*, discussed pitfalls in planning and selecting the catalyst evaluation method, and provided best practices to guide refiners towards an optimal hydroprocessing catalyst selection.

Part 2 discusses critical aspects of a hydroprocessing catalyst testing program and presents readers with best practices/suggestions to ensure a successful benchmarking campaign.

Industrial vs. laboratory-scale fixed-bed reactor. While refiners conduct a catalyst testing campaign to assist their catalyst selection process, experimental aspects remain that require great care and consideration to ensure meaningful results and minimum biases between catalyst vendors.

First, the laboratory-scale fixed-bed reactor should effectively simulate phenomena in the commercial-scale reactor. In most hydroprocessing applications, the trickle flow regime is predominant. Scientists/engineers have proposed experimental techniques to minimize the effects of non-ideal flow behaviors on the performance of a laboratory-scale reactor.[4] For example, a minimum value of reactor-to-particle diameter ratio must be ensured to reduce wall effects. Typically, laboratory-scale fixed-bed reactors have a higher than usual length-to-diameter ratio than commercial reactors for good similarities between two different scales.

The differences between these two different scales are summarized in **TABLE 2.** Even with perfect hydrodynamics, other experimental factors can still cause discrepancies between the laboratory-scale experiment and the commercial process. In the authors' hydroprocessing experience, the laboratory-scale unit's overall performance was often poorer than the commercial one. Other critical experimental aspects are discussed here to better explain how discrepancies can occur, as well as the practical solutions to address them.

LHSV vs. WHSV. Linear hourly space velocity (LHSV) is a widely accepted parameter to connect a laboratory-scale reactor with a commercial one. LHSV is a ratio between volumetric feed flowrate (m^3/hr) and reactor volume (m^3)—as such, the typical unit of LHSV is 1/hr.

TABLE 2. Differences between industrial and laboratory-scale fixed-bed reactor

Parameter	Industrial-scale reactor	Laboratory-scale reactor
Diameter	1 m–6 m	0.2 cm–4 cm
Length	10 m–50 m	0.5 m–2 m
Flow distribution	Requires flow distribution devices to wet the catalyst beds evenly	Using inert material is enough to promote good flow distribution and catalyst wetting
Mode of operation	Non-isothermal: requires interbed cooling or quenching in many cases	Isothermal: no need of interbed cooling, ideal for kinetic study
Axial dispersion	Negligible	Significant in some cases
Catalyst wetting	Very good	Generally poorer
Mass and heat transfer	Very good	Generally poorer
Channeling and wall effects	Negligible	More significant in most cases

Despite its popularity, the authors do not fully agree with using this parameter to simulate the commercial-scale reactor for multiple reasons, especially for scale-down applications (e.g., refinery catalyst benchmarking).

Two primary techniques are available to pack the catalyst inside a commercial-scale reactor: sock loading and dense loading. The catalyst amount for a fixed volume can differ by up to 20% by employing a different loading technique. In other words, the difference in overall activity could be as high as 20%. This discrepancy is particularly crucial for catalysts with extrudate shapes, while the same effect is less pronounced for spherical ones. It is not uncommon to see different catalyst vendors propose a different catalyst loading technique for the same reactor.

Additionally, different catalysts have a unique loading density from different designs and materials.

For these reasons, the authors recommend that every scientist/engineer adhere to weight hourly space velocity [(WHSV), m^3 of feed/hr/kg of catalyst] to improve the similarity between two different scales. Using a specific WHSV for each catalyst vendor is the best option.

Care must be taken which catalyst weight basis is referred to when specifying WHSV (e.g., wet or dry). Modern hydrotreating catalysts usually contain the organic additive (typically organic acid) to promote Type II active sites. Although this organic additive decomposes during the catalyst activation, it contributes 16 wt%–30 wt% of the fresh catalyst weight, depending on the level of metal contents.

H_2 partial pressure and H_2-to-oil ratio. In the authors' experience, typical maximum pressures for laboratory-scale units vary between 160 barg and 175 barg. Despite the use of pure H_2 being prevalent in laboratory-scale experiments [which also diminishes hydrogen sulfide (H_2S) inhibition effects], this pressure limitation still impacts the hydrogenation level, N slip control and the deactivation rate in high-pressure hydrocrackers and residue hydrotreaters (see **TABLE 3** for typical pressure ranges).

Additionally, this limit also affects some product properties from poorer hydrogenation (e.g., cetane number or smoke point). Theoretically, this constraint also impacts the cracking activity, but the effect is only marginal. It is uncommon to see poorer product qualities from a laboratory-scale experiment compared with a commercial unit.

To compensate for lower H_2 partial pressure, the H_2-to-oil ratio can be proportionally increased during the test campaign. For example, if the H_2 partial pressure is off by 10%, the H_2-to-oil ratio should be increased by the same proportion. **Note:** A higher H_2-to-oil ratio cannot fully compensate for lower H_2 partial pressure, but it will surely help.

For most cases, refiners use their actual H_2-to-oil ratio as a target for their pilot plant testing campaign. In the authors' experience, recycle gas flows are notoriously incorrect. The actual gas molecular weight can significantly differ from the one used in the distributed control system (DCS) calculation. This flowmeter does not impact the material balance across the unit; therefore, it is rarely corrected by the refinery. Moreover, the recycle gas sample is seldom taken for units without an amine scrubber due to a high H_2S level; a cartridge of lead acetate is required when taking the sample to lower H_2S concentration to an acceptable level. To calculate a realistic H_2-to-oil ratio, the molecular weight difference must be considered (the uncorrected recycle gas flow reading can be off by up to 20%).

Catalyst loading scheme. Various assumptions have been used for the catalyst loading scheme, such as:

- Replacing the demetallization catalysts with the main hydrotreating catalysts
- Excluding the demetallization catalysts.

The main argument to support these assumptions is that the test campaign is relatively short, therefore demetallization catalysts are not needed (e.g., negligible metal deactivation). This reasoning might be valid if different main hydrotreating catalysts are benchmarked for an R&D program. However, this is not the case for the refinery catalyst benchmarking program, where the ratio between the demetallization catalysts and the main hydrotreating catalysts varies with catalyst vendors.

The catalyst vendor that proposes more demetallization fraction will be given extra activity advantages during the test campaign if the refiner adopts the first assumption. On the contrary, the catalyst vendor with less demetallization fraction will be less affected by excluding the demetallization catalysts if the refiner goes with the second assumption. Some demetallization catalysts have fairly high activity. Excluding them from the test campaign can distort the results.

For example, the catalyst loading scheme with more demetallization fraction tends to require a higher degree of reaction temperature increment than the one with less demetallization fraction, therefore a higher tendency of thermal cracking and crossing aromatics saturation equilibrium.

For these reasons, it is recommended to include both the

TABLE 3. Typical process conditions of various hydrotreating and hydrocracking processes[5]

Application	Petroleum fraction	Reaction phase	LHSV (h^{-1})	H_2 partial pressure, bara	Temperature, °C	H_2/oil (Nm^3/m^3)	H_2 consumption (Nm^3/m^3)
Hydrotreating	Naphtha	Vapor	3–8	10–20	320	60	2–10
	Kerosene	Vapor/Liquid	2–5	20–30	330	80	5–15
	Gasoil	Vapor/Liquid	1.5–4	25–40	340	140	20–40
	VGO	Vapor/Liquid	1–2	50–90	360	210	50–80
	Residue	Vapor/Liquid	0.2–0.5	80–130	370–410	> 525	100–175
Hydrocracking	VGO	Vapor/Liquid	0.5–1.5	90–200	380–430	1,000–2,000	150–300
	Residue	Vapor/Liquid	0.1–0.5	120–210	400–440	1,000–2,000	150–300

demetallization and main hydrotreating catalysts, where applicable, to minimize biases between catalyst loading schemes. Effort must be made to achieve a realistic ratio between the demetallization and main hydrotreating catalysts.

Finally, assumptions should be clearly communicated with all catalyst vendors about your assumptions. In one instance, a major hydrocracking catalyst supplier missed an email about the loading scheme assumptions and suffered from an extremely high pretreating temperature during the catalyst benchmarking campaign. That catalyst supplier even sent its R&D expert to Germany to investigate the issue only to discover that they did not read the email closely enough.

Feedstock. A catalyst supplier once conducted a pilot plant testing campaign for a VGO hydrocracker and included 1% condensate residue (fuel oil blending component). This condensate residue contains a high level of asphaltene, but the refinery nonetheless tried to convert a small amount of it for additional profit. The result is shown in **FIG. 4.** In this case, the reactor inlet is small, and there is no feed filtration system.

Be careful with the choice of feedstocks during the hydroprocessing catalyst testing program. A dirty feed can easily plug a small flow channel and fail the experiment. In this case, the condensate residue merely contributes to product yields, H_2 consumption and product properties. Excluding it from the short-term running campaign would be best, considering the damage it could create to the laboratory-scale system. A filtration system will help, but is no guaranteed solution for dirty feed. For coker oils, maintain positive N_2 pressure from the sample point until the laboratory to prevent subsequent gum formations from olefinic compounds.

In addition to site-specific feedstocks, refiners may further save feed shipment costs by asking the independent laboratory to provide non-specific sulfiding oil for catalyst activation. These independent laboratories usually have leftover sulfiding oil from their previous campaigns.

To better reflect the commercial operation, refiners may also consider adding dimethyl disulfide (DMDS) to simulate the effects of H_2S in the recycle gas.[3] Other species, such as quinoline, can also be added to the feedstock to adjust feed composition.

Recycle operation and recycle cutpoint (RCP). Many hydrocrackers employ a recycle stream to optimize product

FIG. 4. Foulants at the reactor inlet during a hydrocracking pilot plant testing campaign.

yields (FIG. 5, top). The product fractionator separates products from the unconverted oil (UCO) in the commercial unit, then recycles UCO back to the reactor for product yield adjustment. Consequently, UCO gets a second chance to be hydrogenated and converted into products, which usually means product quality improvements.

Unfortunately, most independent laboratories cannot perform a true recycle operation. Instead, a simulated recycle operation is usually achieved by feeding a UCO sample from the actual unit to the pilot-scale reactor and adjusting the single-pass conversion across the pilot-scale reactor to match the reality (FIG. 5, bottom).

Although this simulated recycle operation seems fair on the surface, as every catalyst vendor is under the same assumption, it can still somehow create biases between each catalyst loading scheme. For example, the catalyst system with a better hydrogenation function than the incumbent one (where the UCO sample is taken from) will experience product quality disadvantages, and vice versa for the loading scheme with a poorer hydrogenation function (e.g., free aromatics saturation from the existing unit).

This can also distort the product qualities for different combined feed ratios (CFRs) and overall conversions to certain degrees. CFR is a ratio between the total reactor feed and fresh feed. A higher CFR means more UCO is recycled back to the reactor inlet. In reality, the smoke point and cetane number usually improve when the overall conversion increases (e.g., from better aromatics saturation/ring-opening). Nonetheless, the laboratory experiment gave the opposite results for the cetane number, as seen in FIG. 6. This discrepancy is from the fact that a higher overall conversion operation (98%) has a higher percentage of UCO, and this simulated recycle stream does not have a second chance to be hydrogenated/ring-opened, but is already converted to products during the first pass.

Moreover, the RCP between diesel and UCO is higher in laboratory-scale experiments, with much sharper fractionation (e.g., much less overlap between diesel and UCO). In many commercial units, the separation between diesel and UCO is poor by nature from lower internal reflux at the point of separation. Moreover, the maximum heater duty can also limit the actual RCP.

From FIG. 7, VGO and UCO from the refinery contain around 10%–16% of diesel (defined as a portion with the boiling point below 370°C), while UCO from the pilot plant only has 4%–5% of diesel. This fact implies that the true conversion of a laboratory-scale experiment is lower than that of a commercial-scale operation. In a sense, hydrocracking catalysts are under less conversion stress in a laboratory-scale experiment.

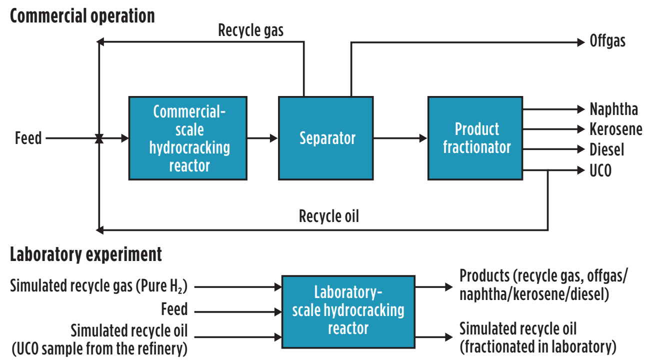


FIG. 5. A single-stage hydrocracker with recycle, commercial operation vs. laboratory.

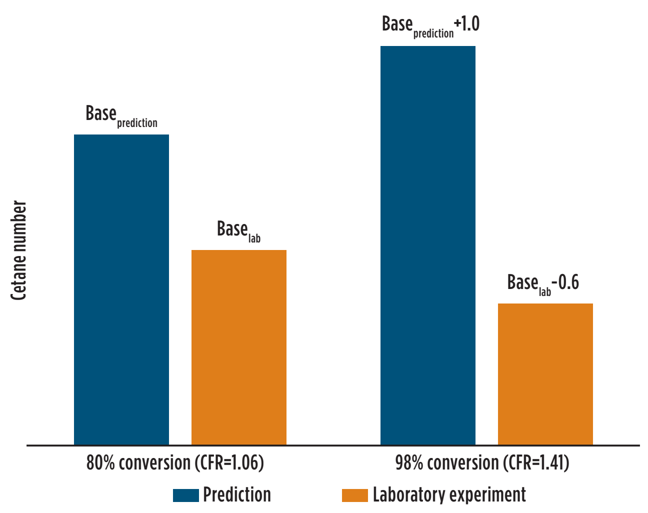


FIG. 6. Cetane number for different overall conversions (and CFRs), prediction vs. laboratory.

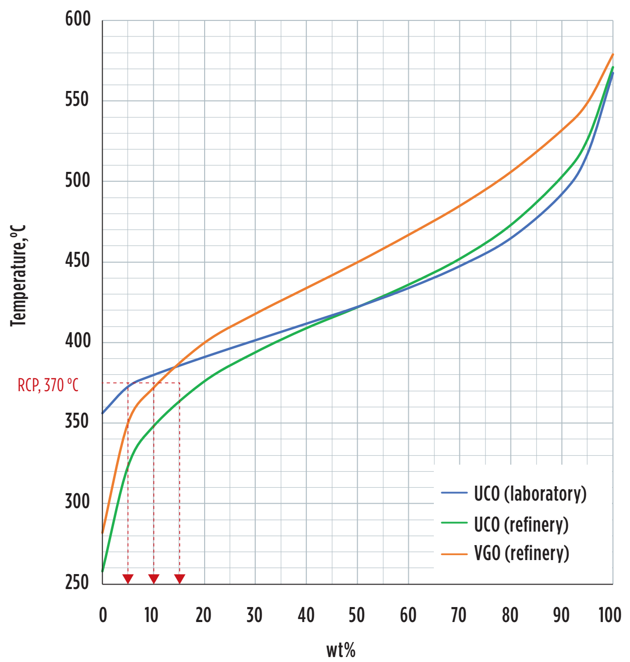


FIG. 7. UCO and VGO distillation by simulated distillation analysis (ASTM D2887), refinery vs. laboratory.

TABLE 4. An example of product cutpoints from ITB

Product	Boiling range, °C
Light naphtha	C_5–90
Heavy naphtha	90–145
Kerosene	145–210
Diesel	210–370
UCO	370–Endpoint

In the authors' experience, these hydrogenation and conversion discrepancies are not that serious. Using oil with lower aromatic contents might help alleviate hydrogenation gaps [e.g., a UCO sample from the start-of-run or gas-to-liquid (GTL) liquid with a similar boiling range]. **Note:** Refiners may consider tailoring the RCP of the hydrocracking catalyst testing campaign to match the operating unit to better reveal the actual performances. Nonetheless, a meaningful comparison between each catalyst loading scheme is still possible despite all mentioned deviations.

Product cut point and properties. Refiners usually specify product cut points between products as guidelines for independent catalyst testing laboratories (**TABLE 4**). However, this does not mean that these product cut points completely agree with other product specifications. For example, the minimum specification of cetane number is 60. During the pilot plant testing campaign, the cetane number will be different for each catalyst vendor despite the same product cut points. To what degree is a better question.

Catalyst vendor A might possess an excellent hydrogenation function and achieve a cetane number of 62, while catalyst vendor B can only offer the cetane number of 58 at the standard product cut points.

To better compare each catalyst vendor, adjusting the cut point between kerosene/diesel to standardize every catalyst vendor is recommended (e.g., target the cetane number of 60). The catalyst vendor with a cetane number > 60 will gain additional diesel from kerosene using this standardization, and vice versa. The wider the gap is, the more diesel gain or loss will be seen. **Note:** The minimum cetane number specification varies with refineries and applications.

A better question is how to standardize the cetane number or other product properties (**FIG. 8**). The most practical solution is to input product yields and properties into the process simulator, then recut each product to match your product specification target.

Depending on the experiment, this product property standardization might or might not change the ranking of each catalyst vendor. For example, suppose the catalyst vendor that is taking the lead already has the highest diesel yield and cetane number with the standard product cut points. In that case, the cetane number standardization will not change the ranking.

Mass balance. Regardless of how proficient the independent catalyst testing laboratory is, there will always be mass balance errors in the real world. In a laboratory-scale experiment, even minor errors can significantly affect the critical parameters (e.g., chemical H_2 consumption). For example, a +/– 2% error is acceptable in the commercial-scale plant, but the same error magnitude could lead to significant errors in elemental balances for laboratory-scale experiments. Ideally, all the key elemental balances, including carbon, H, S and N, should be close to 100% for hydroprocessing applications.

Liquid flow is based on weight measurement for both feeds and products, while gas flow is measured by a flowmeter and an online gas chromatography (GC) analyzer. Typically, gas flowmeters have lower accuracy and are notorious sources of mass balance errors. Additionally, routine activities, including product sampling, sample transfer and product fractionation, can also result in light liquid product losses. As such, measured product flows tend to be less than feed flows, especially in hydrocracking applications (**FIG. 9**). To what degree is a more interesting question.

Typically, the error magnitude increases with processing severity (e.g., catalyst acidity or conversion level). For example, a naphtha selective hydrocracker makes more light products than a diesel selective one, therefore resulting in poorer mass balances from more offgas and naphtha. For a commercial catalyst benchmarking campaign, these mass balance errors can lead to millions of dollars difference in revenue from misleading product yield structure and chemical H_2 consumption.[5]

As illustrated in **FIG. 9,** the pretreating section alone (highlighted in blue) made minimal light ends and is closer to 100 wt% mass balance. For this data set, the error magnitude is less than 0.5 wt% for all data points (exclude the catalyst

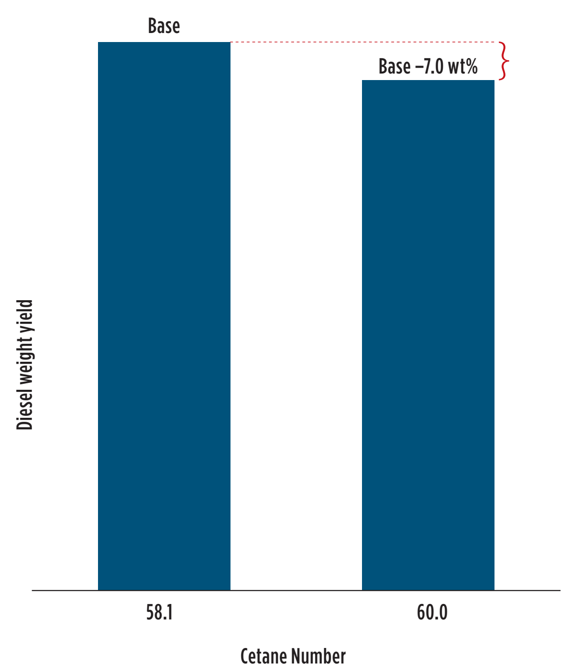


FIG. 8. Diesel yield correction by standardizing cetane number from 58.1 to 60.

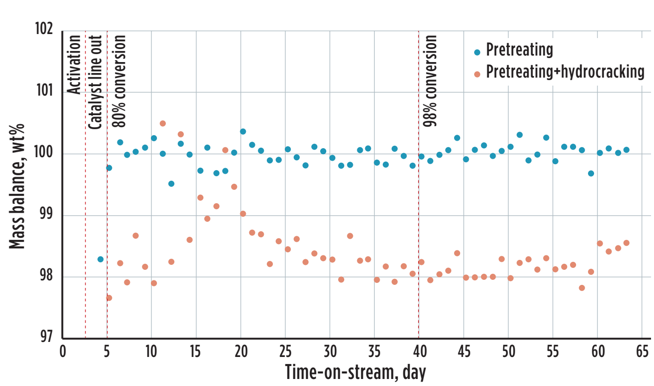


FIG. 9. Overall mass balance vs. time-on-stream of a hydrocracking catalyst benchmarking campaign.

TABLE 5. A demonstration of how data reconciliation could minimize errors for a 54,000-bpd hydrocracker

ΔChemical H_2 consumption, raw-reconciled	80% conversion	98% conversion
wt%	0.12	0.14
Tpd	9.5	11.2
$ MM/cycle	17	20.1

line-out period). Conversely, the hydrocracking section (highlighted in orange) further increased light ends yield and mass balance errors, up to 2.3 wt%.

As a best practice, the authors recommend including a scaling factor into mass balance equations to improve the mass balance of every element. **Note:** The mass balance equations shown below are for demonstration purposes only and are specifically for hydrocracking applications. It is assumed that 100% sulfur and N in the feedstock are converted into H_2S and ammonia (NH_3), respectively, to simplify the mass balance equations.

The overall mass balance equation is shown in Eq. 1:

$$F_{feed} + F_{chemical\ hydrogen} = k_{scaling} \times (F_{gas\ product} + F_{liquid\ product}) + F_{H_2S} + F_{NH_3} \quad (1)$$

The H mass balance equation is shown in Eq. 2:

$$F_{chemical\ hydrogen} = k_{scaling} \times (C_{hydrogen,gas\ product} \times F_{gas\ product} + C_{hydrogen,liquid\ product} \times F_{liquid\ product}) + [(2 \times AM_H)/(MW_{H_2S})] \times F_{H_2S} + [(3 \times AM_H)/(MW_{NH_3})] \times F_{NH_3} - C_{hydrogen,feed} \times F_{feed} \quad (2)$$

The carbon mass balance equation is shown in Eq. 3:

$$C_{carbon,feed} \times F_{feed} = k_{scaling} \times (C_{carbon,gas\ product} \times F_{gas\ product} + C_{carbon,liquid\ product} \times F_{liquid\ product}) \quad (3)$$

Based on the authors' experience, it is impossible to achieve 100% mass balance for every species. Focusing on the H and overall mass balance (e.g., target 100%) for hydroprocessing applications is recommended, as these two mass balances will directly impact chemical H_2 consumption and product yields (**TABLE 5**). As such, the carbon mass balance will be a resultant. In most cases encountered by the authors, the carbon balance nonetheless improved, although it was not perfect.

Where applicable, use multiple data sets of elemental composition (either by EA or NMR) to ensure consistent final ranking results. As a precaution, this data correction practice should be carefully performed to avoid any misleading results from the data correction itself; performing this when the raw mass balance errors exceed 2.0 wt% is not recommended.

Takeaways. Although the economic direction is clear—e.g., maximize middle distillate from a hydrocracker—multiple price sets should always be used to rank hydroprocessing catalysts. This sensitivity analysis ensures the robustness of the final ranking against market fluctuations.

While key ideas presented here are based on hydroprocessing applications, they can be modified/tailored to match other catalyst applications, as well (e.g., elemental balance and data correction technique). HP

NOMENCLATURE

$k_{scaling}$, scaling factor: > 1.00 in case of product loss or < 1.00 in case of product gain
$F_{chemical\ hydrogen}$: mass flowrate of H_2 consumed by chemical reactions or chemical H_2 consumption
$F_{gas\ product}$: mass flowrate of the gas product
$F_{liquid\ product}$: mass flowrate of the liquid product
F_{feed}: mass flowrate of the liquid feed
F_{H2S}: mass flowrate of H_2S, calculated based on the feed flowrate assuming 100% sulfur removal
F_{NH3}: mass flowrate of ammonia, calculated based on the feed flowrate assuming 100% N removal
$C_{hydrogen,\ gas\ product}$: H content of the gas product
$C_{hydrogen\ ,liquid\ product}$: H content of the liquid product
$C_{hydrogen,\ feed}$: H content of the liquid feed
$C_{carbon,\ gas\ product}$: carbon content of the gas product
$C_{carbon,\ liquid\ product}$: carbon content of the liquid product
$C_{carbon,\ feed}$: carbon content of the liquid feed
MW_{H2S}: molecular weight of H_2S, 34.08 g/mol
MW_{NH3}: molecular weight of NH_3, 17.03 g/mol
AM_H: atomic mass of H, 1.01 g/mol

ACKNOWLEDGEMENT

The authors would like to thank Richard Palluzi from Richard Palluzi LLC, Shaun Dyke from PetroQuantum, Tiago Vilela from Avantium, Tino Badra (Independent Consultant) and Meng Loong Chua from Shell Catalysts & Technologies for helpful comments.

LITERATURE CITED

1. Trotus, I.-T., J.-C. Adelbrecht, F. Huber, N. Pongboot and T. Upienpong, "Advances in catalyst testing," *PTQ*, Q4 2019, Digital Refining Processing, Operations and Maintenance, online: https://www.digitalrefining.com/article/1002388/advances-incatalyst-testing#.YYl5dmBBxPZ
2. Vilela, T., "Refinery catalyst testing," *PTQ*, Q2 2019, Digital Refining Processing, Operations and Maintenance, online: https://www.digitalrefining.com/article/1002336/refinery-catalyst-testing#.YYl5m2BBxPY
3. Davy, L., J. Castro, R. Wessels, M. Rey-Bayle, I. Merdrignac and B. Celse, "Global methodology for catalyst screening and optimization process—Application to mild hydrocracking," *Industrial & Engineering Chemistry Research*, 2020, online: https://pubs.acs.org/doi/10.1021/acs.iecr.0c04120
4. Mederos, F. S., J. Ancheyta and J. Chen, "Review on criteria to ensure ideal behaviors in trickle-bed reactors," *Applied Catalysis A*, 2009, online: https://www.sciencedirect.com/science/article/abs/pii/S0926860X08007059
5. Ancheyta, J., *Modeling of processes and reactors for upgrading of heavy petroleum*, 1st Ed., Routledge Taylor & Francis Group, CRC Press, 2013.

NATTAPONG PONGBOOT is an experienced Chemical Engineer with broad experience in refining and petrochemical technologies. He has extensive experience in hydroprocessing catalyst selection and testing as both a refiner and licensor. Before partnering with Global R&D, he held various technical roles with Honeywell UOP, PTT Global Chemical and SCG Chemicals, with broad exposure to process and equipment design, refinery operation, plant troubleshooting and process optimization. Additionally, he is also a technical instructor for www.chemengedu.com. Mr. Pongboot obtained B.Eng. and M.Eng. degrees in chemical engineering from Kasetsart University. The author can be reached at nat.pongboot@gmail.com or nattapong.globalrd@gmail.com.

THANAWAT UPIENPONG is a Senior Refinery Process Engineer at PTT Global Chemical Plc. based in Rayong, Thailand. He is a chemical engineer with more than 9 yr of experience in the refining industry. He specializes in refinery operation, troubleshooting, optimization, startup/shutdown and catalyst selection with broad exposure to various refining processes, such as thermal conversion, diesel hydrotreating and hydrocracking. Mr. Upienpong earned his M.Eng. degree in chemical engineering from King Mongkut's University of Technology Thonburi in Thailand. The author can be reached at Thanawat.U@pttgcgroup.com or tai. thanawat@gmail.com.

WIRINYA KARUNKEYOON is a Refinery Process Engineer with several years of experience in water treatment, power generation and diesel hydrotreating. She is affiliated with PTT Global Chemical (Site 6: Refinery) in Thailand. Ms. Karunkeyoon obtained her B.Eng. degree in chemical engineering from King Mongkut's University of Technology North Bangkok, and her M.Eng. degree in chemical engineering from King Mongkut's University of Technology Thonburi. The author can be reached at Wirinya.K@pttgcgroup.com.

C. RENTSCHLER, Al Bassami Industries, Lititz, Pennsylvania; **A. PARMAR,** MilliporeSigma, a Merck KGaA Company, Sheboygan, Wisconsin; and **G. SHAHANI,** Independent Consultant, Allentown, Pennsylvania

Overcome inflation and supply chain issues in construction

Inflation and supply chain disruptions are not new to the construction industry, but the severe combination of these factors and the rapid pace of change driven by the COVID-19 pandemic are unprecedented. This disruptive change in essential factors affecting construction company performance and sustainability requires companies to make quick adjustments to project execution. While these and other challenging issues have been spiking quickly due to COVID, the problem is not short term and will require focused corrective efforts for the near future.

The major contributors to the latest construction industry woes are material shortages and spiking prices when materials can be found. These challenges are further compounded by a tougher labor market. Many domestic mills and fabrication facilities were forced to shut down when the pandemic hit. When production facilities were allowed to reopen, they faced issues in receiving their raw materials, establishing sufficient labor and securing reliable shipping. All of this led to material availability issues and higher material costs. Foreign manufactured products and materials also faced production and shipping shutdowns during the pandemic, and this further exacerbated an already grave situation regarding supply shortages.

The combination of spiking commodity prices, supply interruptions and tight labor markets—mainly from the pandemic—has been fueling market uncertainty and inflation. In some cases, the escalating cost of equipment, materials and labor are making projects financially unviable. In other cases, projects are being postponed. Projects that were already under execution when the pandemic hit were impacted both financially and schedule-wise. Future viable projects will be impacted by increased project demands spurred by low interest rates causing price pressures and supply shortages. Solving the problem is not a matter of just raising prices, since construction companies must still be competitive to win much-needed work. Contractors and owners must take steps to deal with the situation as best they can by addressing those factors within their means. Suggested steps are discussed in this article.

More than anything, the current situation driven by the pandemic should not be looked at as a once-and-done scenario. History has shown that supply shortages and increasing inflation can be sporadic events. While experience has demonstrated that the price spikes eventually subsided (in nearly all cases), contractors that were unable to pass on the price increases for an extended period were harmed.

Contractors should be equipped with the knowledge and structure to fight through the inflation situations when they occur. An example is to reexamine the terms and conditions of contracts, especially if they are based on fixed or lump sum pricing and/or completion deadlines. It is a good idea to review sourcing and procurement strategies for materials and hiring plans to expand the resource pool. Lessons learned from the current pandemic situation are a perfect knowledge platform for situations that may occur in the future.

Communication between owners and contractors is key during times of inflationary material costs. Owners should expect different pricing and may want to consider design variations, cost sharing and time flexibility. Contractors must closely monitor costs and delivery schedules for materials and communicate this information to the owner both during the bid stage and throughout construction. In the end, price spikes do eventually reverse course—when this happens, both the owner and contractor benefit. Along the way, staying aligned will promote a cooperative working relationship, avoid dissension and lead to a successful project, even during challenging times.

Commodity prices and availability. Media has been rife with references to inflation on a global basis caused by supply-demand imbalances and easy monetary policies.

The Associated General Contractors of America(AGC) published an alert in April 2022 regarding how inflation is impacting the construction industry.[1] The trends presented here apply to non-resi-

dential construction. Additional details can be found in literature[1] and Bureau of Labor Statistics. FIG. 1 shows that while the material costs went up by 21% in the 12 mos prior to February 2022, bid prices increased only 17% during the same period. This illustrates the contractor's ability to pass on some price increases to owners.

As a result, contractors have to look for creative solutions to deal with this issue. Based on prior experience, this issue will not be transitory and is likely to persist for several years until equilibrium returns to the marketplace.

Project development. Project development usually requires a detailed financial analysis to obtain approval for investment. This entails a discounted cash flow analysis, which requires an estimation of revenue, capital and operating costs and risk. If a project has an internal rate of return (IRR) that is higher than the cost of capital—accounting for risk—the project will be approved.

Every component of future revenue and costs contains uncertainty; future revenue depends on volumes and price; operating costs depend on unit consumption and unit costs of raw materials, utilities and labor; and capital costs depend on the cost of engineering, procurement of major equipment and the cost of construction. While operating costs can be accurately estimated with modern process simulation software, capital costs are much more difficult to estimate precisely due to the variability of labor and materials. As a result, many owners and their engineers carry a contingency in their capital cost estimation to cover unexpected costs.

For large projects like the cracking furnace shown in FIG. 2, construction can be ~30% of the total installed cost. If the capital cost increases, the expected returns from a project can jeopardize the economic viability of a project. Over a period, if enough projects are not approved, the aggregate demand for construction materials and labor will decline. This will help moderate pricing, and eventually prices will come down.

However, in the short term, creative solutions are needed to deal with uncertainty. It must be noted that all the projects that were once put on hold or postponed may emerge again in the future and will end up creating a supply-demand issue. It is prudent to continue seamless communication with the owners (or potential owners) during this postponement period impacted by the pandemic. The same applies with the current or potential suppliers of materials used in construction. Lead times for equipment and a large range of construction materials have grown exponentially due to the current situation.

Coping with uncertainty. While risk and uncertainty cannot be eliminated, ways to manage risk are discussed below. All projects have a certain degree of risks in executing—it is always prudent to identify risks and make attempts to mitigate the identified risks. A risk register should be maintained in agreement with the customer, and if needed, with customer inputs incorporated. These registers would address any uncertainties, such as pricing, the management of change (MoC) process and its impact on the project, and safety aspects.[2,3]

Schedule flexibility. The ongoing COVID pandemic over the last 2 yr has made the world a more uncertain, volatile and a complex place to continue business. Contractors should explore opportunities to continuously evolve to support business, such as exploring flexibility in the project schedule if it allows mitigating risks, especially escalating costs. Allowing reasonable time during the beginning phase of a project (e.g., in the engineering phase) facilitates more value creation or addition in the front end. This may result in a reduction in bill of materials or promote alternative materials to reduce costs, gain process efficiencies, reduce eventual operating costs, encourage creative automation, etc. Possibilities to reduce overall costs can be found, especially if timing is not a critical factor due to current business conditions. An overall review of the project schedule makes complete sense to optimize cost factors. If possible, tweaks in a project schedule should be made to accommodate design reviews to reduce project cost, accommodate expected lead-time issues, and synchronize/minimize staffing and workforce deployment in different phases of a project, especially in construction.

Open book with customer. Transparency shown by the contractor leads to a more confident and trusting owner, and thereby results in a closer working relationship. Disclosing a certain level of information—with a non-disclosure agreement in place, if required—helps gain an owner's trust. At the same time, assisting the owner to ensure the degree of change management also helps. Depending on the project phase and execution model adopted, change orders can be expected in real life. This is also the basis for contingencies defined by the owners.

Long-term relationship with vendors. A returning customer with an invitation for bid tells a lot about past performance. The relationship with a customer should nurture itself with con-

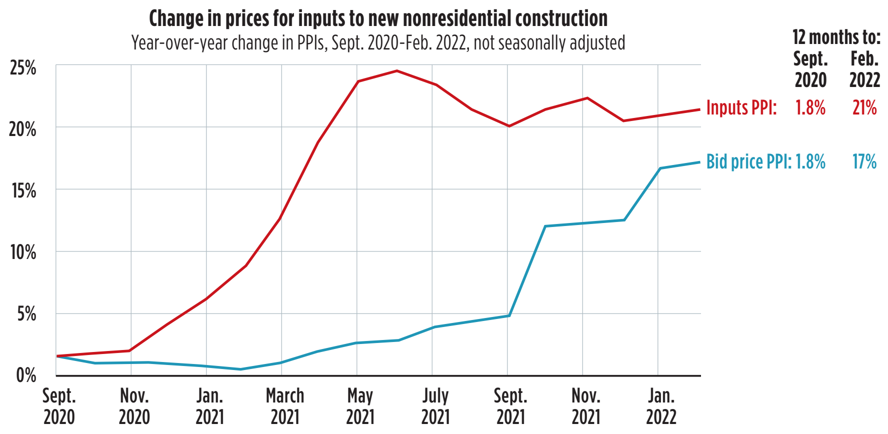


FIG. 1. Change to construction inputs and bid prices, September 2020–February 2022. Source: Associated General Contractors of America and Bureau of Labor Statistics.

FIG. 2. Large cracking furnace in the Middle East. Source: Al Bassami Industries.

sistently good performance and quality of work, agreed-upon timeliness and cost factors. With a carefully maintained long-term collaboration, partners will have a stake in working through inflation and supply chain issues as a team. Rather than finger pointing, there will be an understanding of the difficulties and a team effort to move the project forward with minimal impact.

Locked-in prices. With any project, an effort should be made to finalize engineering as soon as possible so material prices can be locked in early. When executing multiple projects with one or more customers, it makes sense to review the common bill of materials to obtain good bulk purchase prices. The savings can be shared with the customer, keeping in mind a long-term relationship. Also, some synergies and efficiency gains can be achieved by deploying labor to project jobs that are similar in nature or short-term repetitive tasks. A project agreement with the customer, whether on a lump-sum or cost-plus basis, allows any foreseeable cost savings to be shared with the customer, further opening the transparency part of the relationship.

Multiple suppliers. A good procurement policy and the team behind it should work creatively to address materials requirements. The availability of materials during the pandemic, especially the longer lead times, poses challenges. This can be mitigated by exploring the availability with multiple suppliers. Eventually, a supplier's relationship can be maintained just like the customer relationship. Strong supplier performance is key to good project execution and a satisfied customer.

Contract with escalation. Contracts with an escalation clause can be agreed with customers and should be mutually favorable and win-win for both parties. A due diligence process should be in place, also adding to transparency.

Takeaways.

Inflation and supply chain concerns can be project development and execution issues that are often ignored as a project progresses—they can become significant problems when they develop. The pandemic has demonstrated the seriousness of these issues as they relate to successful project execution. As a result of the pandemic, many companies have been faced with a large disparity between their bid prices and actual costs, with no ability to pass on increases. Failure to take the proper steps to address these issues can have a dramatic effect on a company's balance sheet.

Maintaining transparency is a good start to a guarded position relative to inflation and supply chain issues. This includes establishing and maintaining a long-term relationship with vendors and keeping an open book with the customer. Buying in bulk and locking in prices early are other positive steps. Maintaining multiple suppliers is a safeguard to inflation and delivery issues, as not all vendors are impacted to the same degree. Having the foresight to structure contracts with escalation is the ultimate protection for inflation, but this may not always be possible in a bid situation.

Finally, open communication between owner and contractor is the best means to work through inflation and supply chain issues. The best approach is always one that ensures cooperation and understanding during project challenges, and the sharing of rewards when times are good. HP

LITERATURE CITED

Complete literature cited available online at www.HydrocarbonProcessing.com.

CARL RENTSCHLER is the U.S. Sales Representative for Al Bassami Industries, a premier global steel fabrication company based in Jeddah, Saudi Arabia. He has more than 40 yr of varied engineering and management experience with Parsons, Air Products and Linde in the power and petrochemical fields. Mr. Rentschler is a licensed Professional Engineer, holds a BS degree in civil engineering from Penn State University, and earned MEng degree from Cornell University.

ASHIM PARMAR is Head of project engineering at MilliporeSigma, a Merck KGaA Company. He brings 28 yr of experience in project management, operations and maintenance, reliability, and green field startups. His primary responsibility is leading projects from conceptual phase to successful completion. Mr. Parmar is focused on continuous improvement and is passionate about the Internet of Things (IoT), automation and digitization with a value-added approach. He has international experience and previously worked for Eastman Chemicals, International Flavors and Fragrances, Symrise Asia Pacific, Novartis and Pfizer Pharmaceuticals.

GOUTAM SHAHANI works as a Sales and Marketing Consultant in the construction industry. He teaches business and mathematics at area colleges, and has more than 40 yr of experience in industrial marketing, business development and asset management at Air Products, Linde and ShureLine Construction. Mr. Shahani has more than 50 publications and patents in the energy and environmental sectors, and holds BS and MS degrees in chemical engineering, as well as an MBA.

G. W. PARK, Kuwait Oil Co., Ahmadi, Kuwait

Prevent corrosion failures at each stage of design, construction and normal operation

Each year, significant economic losses due to corrosion failures occur in the refining and petrochemical industries; unfortunately, many of these losses could have been prevented. Petroleum refining today is unusually sophisticated compared to the single-shell stills of the 1800s, and the industry shows every indication of becoming even more complex.

Chemical and mechanical engineering advances are increasing product yields and improving plant operating reliability, and crude units have been highly integrated with secondary and petrochemical plants, such as fluidized catalytic cracking units (FCCUs), hydrocracking units, and delayed coker and polymer plants. These units are operating under various corrosive materials and operating conditions (FIG. 1).

One result of these changes is a growing emphasis on materials engineering and increased interest in high-alloy, corrosion-resistant steels—such as stainless steels, nickel alloys, copper alloys and titanium alloys—to cope with a wide variety of raw crudes, oil intermediates and

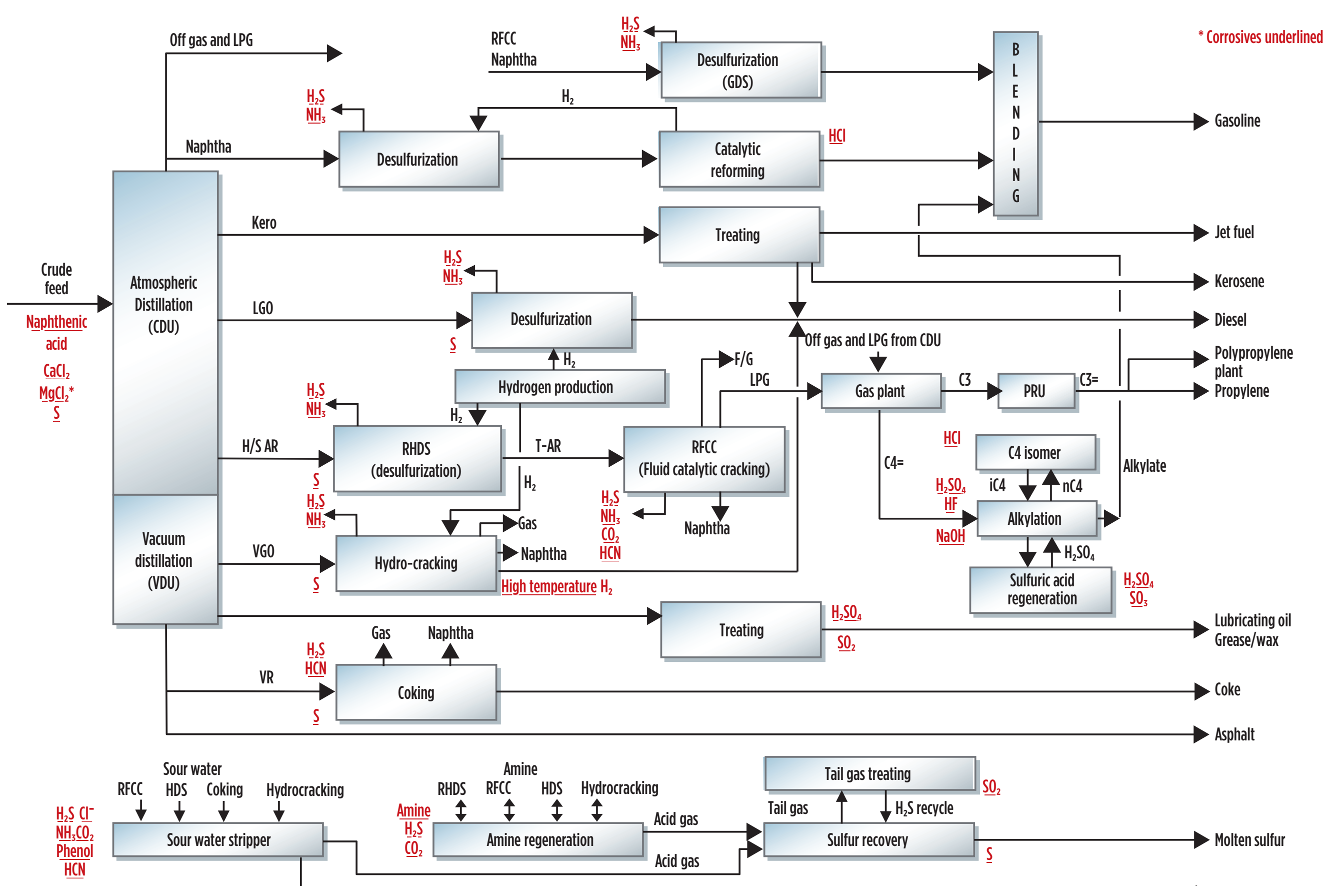


FIG. 1. Typical block diagram of a petroleum refinery with corrosives.

products. It is obvious that substantial savings can be obtained in many types of plants through the use of corrosion-resistant materials of construction.[1] To do this, close cooperation between the inspection (corrosion) engineer and process and design personnel is required before a plant is built to eliminate or reduce maintenance costs in many cases.

Slight changes in the process sometimes reduces the corrosiveness of plant liquors without affecting the process itself, permitting the use of less expensive materials. These changes can often be made after the plant is in operation, but original preventive measures are more desirable.

Sometimes, design errors can occur during communication among process licensors, foreign engineering companies, local engineering companies and equipment vendors. This article presents several case histories of refineries where inspection and investigation identified potential and actual clues of corrosion failures caused by improper internal manufacturing by a tower vendor, an incorrect design of a piping spec break by an engineering company, inappropriate material selection by license companies and a lack of understanding of chemical properties. Additionally, this article focuses on how process engineers can prevent corrosion failures at each stage of design, construction and normal operation, based on the author's experience.

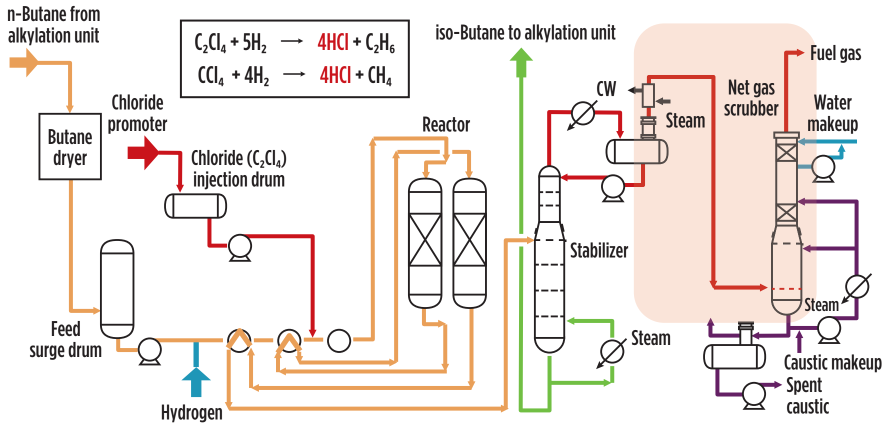


FIG. 2. C_4 isomerization process.

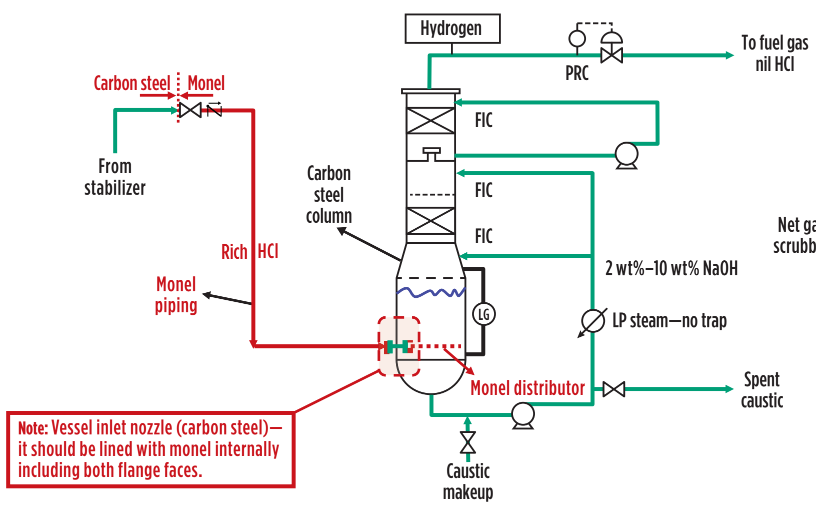


FIG. 3. Monel lining missing at the inlet nozzle to the net gas scrubber.

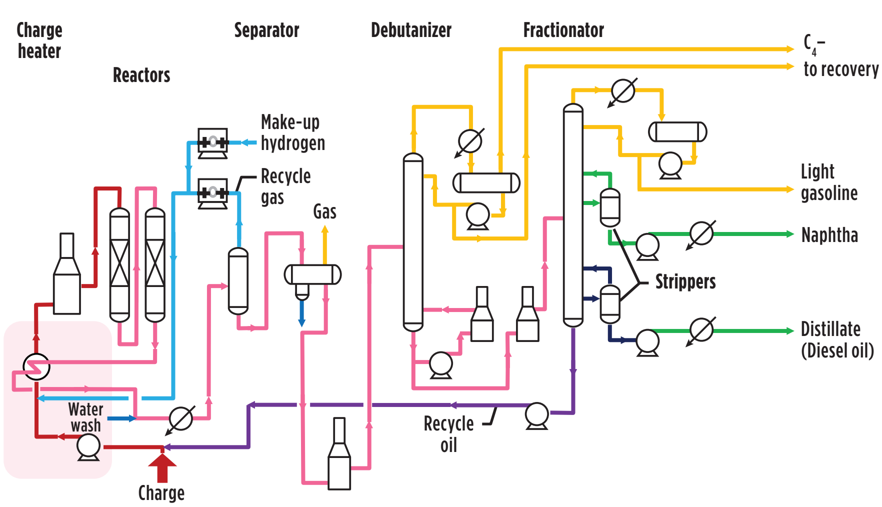


FIG. 4. Hydrocracking process.

CASE 1: MONEL LINING MISSING AT THE INLET NOZZLE TO THE NET GAS SCRUBBER OF A C_4 ISOMER

The C_4 isomer (butamer process) is specifically designed for the catalytic isomerization of butanes on platinum type catalyst with the chloride promoter, which decomposes to hydrogen chloride (HCl, corrosives) downstream of the chloride injection point. The typical process schematic is shown in **FIG. 2.** The principal reactions of the isomerization are shown in Eqs. 1 and 2:

$$CH_3\text{-}CH_2\text{-}CH_2\text{-}CH_3 \rightarrow \begin{array}{c} CH_3 \\ | \\ CH_3\text{-}CH\text{-}CH_3 \end{array} \qquad (1)$$

$$C_2Cl_4 + 5H_2 \rightarrow 4HCl + C_2H_6 \quad (\text{Temp.} > 105°C) \qquad (2)$$

This decomposed HCl will meet the aqueous solution downstream of the stabilizer overhead system (where very corrosive material forms) and then is neutralized in the net gas scrubber by caustic,

sodium hydroxide (NaOH). Therefore, the piping from the stabilizer overhead to the net gas scrubber is made of monel material [66%–67% nickel (Ni), 29%–30% copper (Cu)], a highly resistant material for HCI aqueous solution that is also applied to the upper part of the atmospheric distillation column.

Problem. According to the contract, a process license company dispatched an inspection engineer to check the vessels and columns in the process just after completion of construction and internal installation. The engineer's background was mechanical; all equipment was inspected, including the net gas scrubber internals, and everything was found to be in working order. As a final cross check, the author performed an inspection with the license vessel drawings following the license inspector's report. A critical inconsistency was found at the inlet nozzle of the net gas scrubber: monel lining was missing inside of the nozzle and flange faces. The licensor specified the note on its column drawing presented in **FIG. 3.**

Cause and solution. This finding was promptly shared with the licensor, and was also compared with the vendor drawings from the manufacturer and the licensor. Consequently, the vendor had not reflected this critical note on its drawings and made

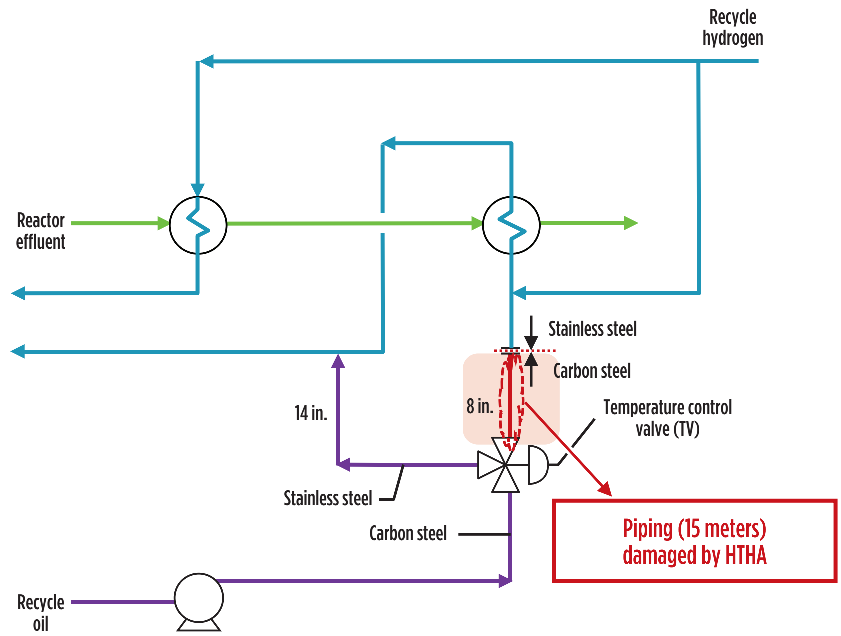


FIG. 5. Piping schematic for the explosion area in the hydrocracking process.

the inlet nozzle without monel lining. The process licensor confirmed that it should be corrected accordingly based on its specification. The modification—monel lining on the inlet nozzle—was performed in the vendor shop near the plant site.

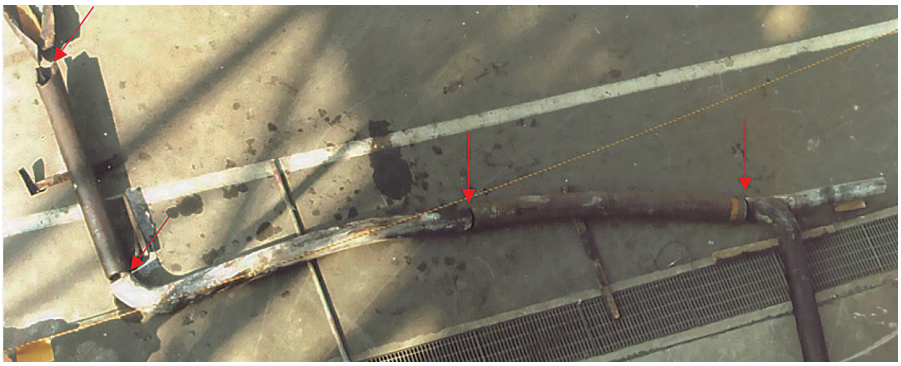


FIG. 6. The ruptured part of the damaged piping in the hydrocracking process.

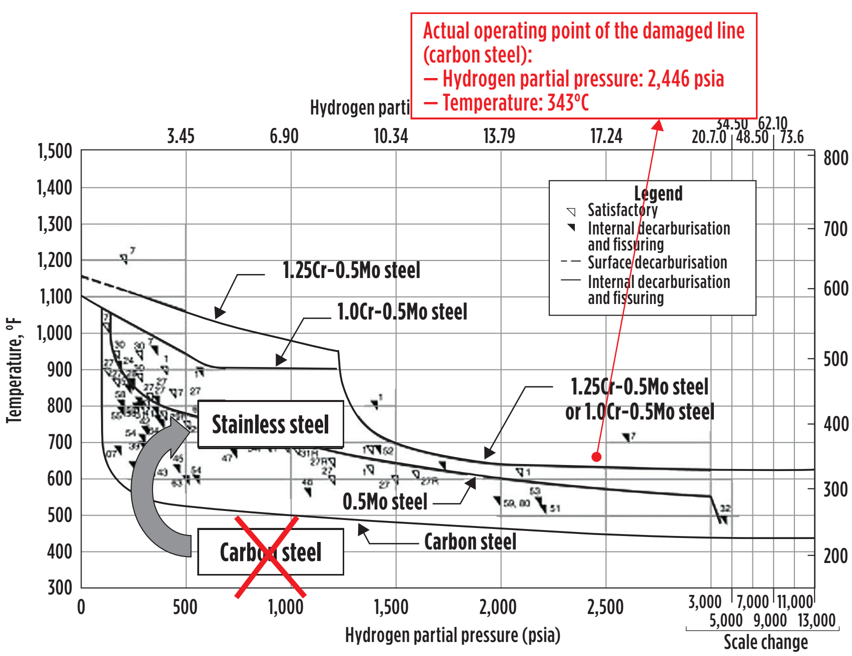


FIG. 7. Nelson chart (API RP941) with the actual operating point.

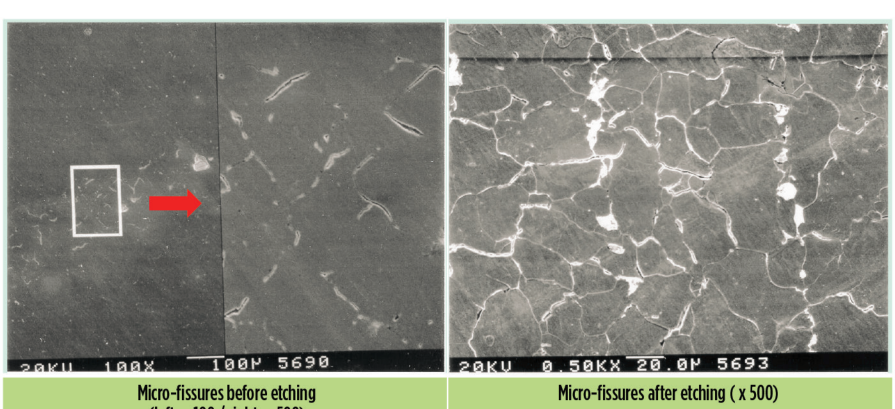


FIG. 8. Micro-fissures of the damaged piping sample due to hydrogen attack (HTHA).

Lessons learned. Most new processes are designed by license companies and then developed by basic and detail engineering companies. Equipment vendors also issue their drawings; process and design engineers should carefully compare them with the originals from process licensors because incorrect interpretations and communication can happen. Secondly, it is highly recommended that process engineers perform an inspection from the point of view of a process.

The above finding was also conducted by an owner-side process engineer after the error by the license mechanical engineer. Since the process expert fully understood that the stream—an HCI aqueous solution—was very corrosive, both sides of the nozzle, upstream piping and monel distributor are made with monel material prior to neutralization with the caustic.

However, the nozzle should be made of carbon steel because the column base material is carbon steel. Therefore, the nozzle should be lined with highly resistant material to avoid acid corrosion. Because this C_4 isomer is part of the FCC plant, significant losses during an emergency shutdown could have occurred within several months of initial operation without this proper and meticulous inspection. To achieve trouble-free operation, process engineers and inspection personnel must conduct in-depth inspections.

CASE 2: EXPLOSION DUE TO HIGH-TEMPERATURE HYDROGEN ATTACK (HTHA) IN THE RECYCLE OIL PIPING OF THE HYDROCRACKING PROCESS

The hydrocracking process is designed to convert petroleum distillates of various boiling ranges into lighter, more valuable low-sulfur fractions. The product will normally be a material that may be fractionated into cuts ranging from naphtha to heavy diesel oil. One typical reaction system consists of a two-stage series reactor system with hydrogen recycle. The first reactor is loaded with a catalyst designed to clean up the charge, allowing the removal of the feed nitrogen and sulfur as ammonia (NH_3) and hydrogen sulfide (H_2S), and to selectively hydrocrack heavier hydrocarbons into

lower boiling point materials. The second stage is employed to further crack the first-stage effluent into the desired boiling range products. Reactor effluent passes through feed/effluent exchangers and product condensers into a high-pressure separator where hydrogen recycle is taken out. The net liquid is then flashed at a lower pressure and temperature before being debutanized and fractionated into the desired products.[2] The typical process schematic is shown in **FIG. 4.**

Problem. In one incident, the recycle oil piping (8 in., 15 m) of the hydrocracking process was ruptured and caused an explosion and large fire. The ruptured portion of the piping is depicted in **FIG. 5** because it is not detailed properly in the overall process schematic (**FIG. 4**). When the inspection team and the author arrived at the site, the ruptured parts of the damaged piping were cut as if with a sharp knife (**FIG. 6**). The burging (swelling) part of the left end piping seemed to be caused by fire and explosion (secondary damage) rather than by the original damage mechanism.

Cause analysis. The material group and the inspection team began investigating the cause of the accident. The operating condition of the damaged piping was checked with the API RP941 Nelson chart, since the carbon steel line was exposed in the hydrogen environment. The chart indicates the material's service limit to avoid decarburization and fissuring, which is used in the high-temperature hydrogen environment. Operating personnel confirmed that hydrogen reverse flow through this carbon steel piping could be formed during turndown operation and startup due to the temporary stop of the recycle oil pump. The actual operating conditions and design values of the line are shown in **TABLE 1.**

> Substantial savings can be obtained using corrosion-resistant construction materials. Close cooperation between the inspection (corrosion) engineer and process and design personnel is required before a plant is built to eliminate or reduce maintenance costs in many cases.

Considering 95% of the hydrogen content in the recycle hydrogen, the actual operating point was marked on the Nelson chart shown in **FIG. 7.** The point exceeded the limit of carbon steel and low-alloy materials, which are susceptible to high-temperature hydrogen attack (HTHA). Stainless steel should have been selected for this application. As described in **FIG. 5,** the 14-in. bypass line from the temperature control valve (TV) was designed as stainless steel, but the section of the 8-in. main line downstream of the TV was made with carbon steel, which showed inconsistency in design since both lines are exposed to high-temperature hydrogen.

In equipment that is exposed to high-temperature and high-pressure hydrogen gas of > 200°C, hydrogen creates methane by permeating into the steel inside and reacting with dissolved carbon or metal carbide in dislocation, inclusion, gross discontinuity, lamination or other inside voids, according to the reaction in Eq. 3:

$$Fe_3C + 2H_2 \rightarrow 3Fe + CH_4 \quad \text{(decarburization reaction)} \tag{3}$$

Because a molecule of methane is large, it is hard to diffuse. The methane pressure increases by accumulating in the grain boundary or the boundary between the grain and nonmetal inclusion. Decarburization then progresses continuously, creating bubbles and fissures (minute cracks) in the steel inside. There is no change in the thickness of metal, but micro cracks are connected and there is lower tensile strength and ductility by decarburization and fissuring. As a result, hydrogen attack occurs. To confirm the cause of this accident, a replica test of metal microstructure was performed by

TABLE 1. Actual operating conditions and design values of the line

	Design case	Operating case
Pressure	260 kg/cm²,g	180 kg/cm²,g
Temperature	370°C	343°C

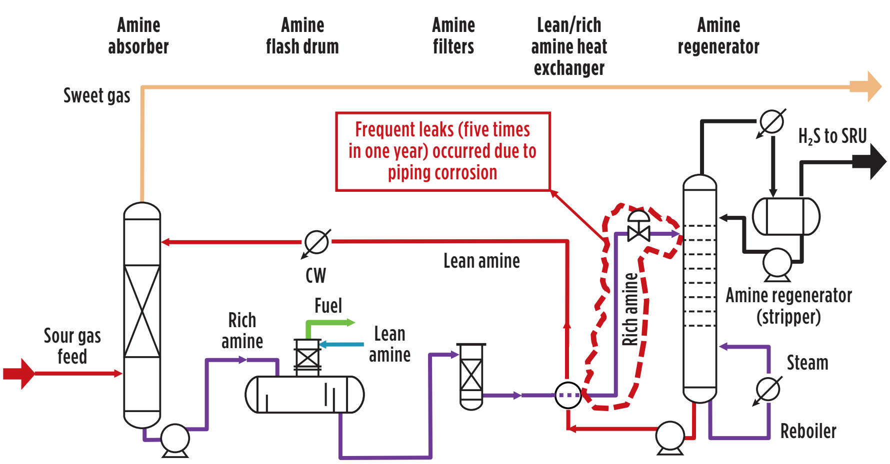


FIG. 9. Amine system: Amine absorber and amine regeneration unit (ARU).

FIG. 10. Amine corrosion in the rich amine piping.

the materials group with the sample of the damaged piping. Numerous micro-fissures were found (FIG. 8).

Additionally, as hydrogen atoms are located in a high-stress area, hydrogen attack occurs easily because it diffuses well. So, a heat affected zone (HAZ) is weaker against HTHA than base metal or weld metal, and decarburized areas and fissuring are often found there. This explains why the line was sharply cut at the points of the piping's HAZ.

Solution. The refinery owner claimed inappropriate piping design and an incorrect piping spec break against the engineering company. Consequently, the damaged piping and 10 m of the recycle oil line upstream of the TV were planned to be replaced with stainless steel material. The installation of check valves to prevent reverse hydrogen flow was also considered.

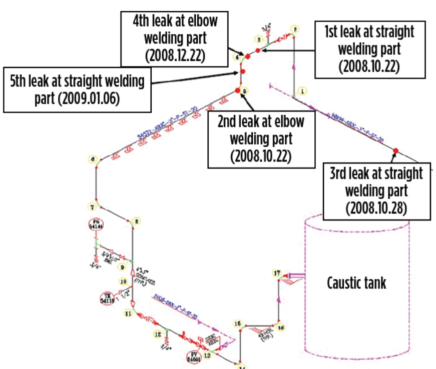


FIG. 11. The piping layout and leak points around the caustic tank.

Lessons learned. To clearly identify the damage mechanism of the corrosion incident, cooperation among concerned teams (operations, process technical services, inspection, R&D, etc.) is vital. Additionally, corrosion and material experts with sound knowledge and experience of various aspects of corrosion, as well as inspection tools and methodologies, are required. For process engineers, a spec break review of piping and equipment is critical by those who are aware of process stream operating conditions and stream components. After a joint review with corrosion engineers, the proper materials for piping and equipment can be selected.

CASE 3: REPEATED CORROSION IN THE PIPING OF THE AMINE REGENERATION UNIT (ARU)

An amine system removes H_2S and carbon dioxide (CO_2) from various fuel gases produced in a refinery to prevent air pollution when the gases are burned. Amine plants consist of two units (FIG. 9): the absorber in which the unwanted gas constituent is removed, and a regenerator in which the amine solution is reactivated. In the absorber, gas flows up through the column and the amine cascades down, providing intimate contact. Sweet gas exits the top, and rich amine leaves through the bottom. In the regenerator, H_2S and CO_2 are expelled by heat and stripping steam. The acid gases go to sulfur recovery units (SRUs) and lean amine solution is returned to the amine absorber.

Problem. In an amine unit, operation and maintenance personnel noticed five leaks in one year in the piping circuit—oddly, the leaks occurred only in the rich amine line between the lean/rich amine heat exchanger and the amine regenerator (within the dotted red line in FIG. 9). The piping material was carbon steel while the tube material of the amine heat exchanger was stainless steel.

Cause analysis. To determine the cause of the leaks, an inspection team investigated the inside of the piping after cutting the line. Internally, severe pitting corrosion with uniform thickness loss was found in the rich amine piping (FIG. 10). This is a typical type of amine corrosion. In this case, the proper material is austenitic stainless steel or duplex stainless steel; the corroded piping was made with carbon steel, which is weaker.

Furthermore, a temperature rise and the current speed are accelerating factors for this type of corrosion, particularly when turbulence is created by acid gas (H_2S and CO_2) flashing in rich amine due to a temperature rise (from 52°C to 100°C) through the tube side of the lean/rich amine heat exchanger and a pressure drop across the flow control valve at the inlet of the amine regenerator. Erosion corrosion can be also shown in this area. As a reference check, the material of rich amine in other amine units was investigated. The rich amine piping downstream of the heat exchangers in other amine units was made with stainless steel material. So, the investigation teams concluded that this repeated corrosion occurred due to incorrect material selection.

Solution. The rich amine piping after the lean/rich amine heat exchanger was replaced with stainless steel material.

Lessons learned. This amine unit had been designed by a local engineering company that had insufficient experience with amine processes. The owners or engineering companies should have consulted with major license companies or major engineering companies. Referencing piping and instrumentation diagrams (P&IDs) for other amine units is also an effective and preventive measure.

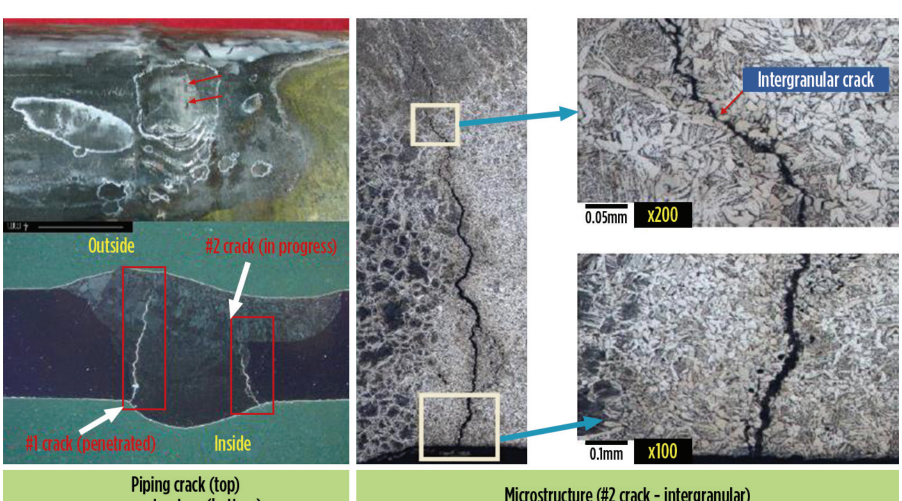


FIG. 12. Sample analysis of the damaged caustic line.

CASE 4: REPEATED PIPING CRACKS OF CAUSTIC MAKEUP LINES DUE TO CAUSTIC STRESS CORROSION CRACKING (SCC)

Caustic (NaOH) is used widely in refining and petrochemical units for the neutralization of acid constituent and pH control of cooling water or boiler feed water. However, it can cause SCC to carbon steel welds, according to the concentration of caustic and temperature (refer to API RP571: NACE caustic soda chart). Stress-relieved carbon steel or nickel alloys can be more resistant to caustic SCC.

Problem. At the makeup line of fresh caustic to the caustic tank in the alkylation unit, five leaks occurred over a period of 3 mos (**FIG. 11**). The service of the 3-in. piping was 25 wt% fresh caustic and it was installed with steam tracing. This caustic was used for neutralization at the caustic wash drum and for acid gas scrubbing at the blowdown vapor scrubber in the alkylation unit.

Cause analysis. An inspection team collected the piping sample at a fifth leak welding part. A crack was visually detected at the circumferential direction of the welding part and several white stains were noticed on the surface of piping due to caustic leak (**FIG. 12**, left top). Through a radiography test (RT), it was determined that the crack penetrated the weld metal beginning at the weld toe (**FIG. 12,** left bottom); its microstructure was investigated after cutting the sample piping.

Intergranular cracking was observed along grain boundaries (**FIG. 12,** right). Intergranular stress-corrosion cracking is typically shown in the alkaline SCC, such as caustic SCC and amine SCC. Intermittent caustic supply to the tank was required about once a week, while steam tracing on the piping was under continuous service. The actual operating temperature of the makeup line fluctuated 30°C–75°C, according to the temperature gauge, while the normal temperature was shown as 38°C on the document. At 38°C, caustic SCC cannot exist; therefore, it was possible that the piping temperature rose due to continuous heat flux from steam tracing, causing the increase of concentration of the remaining caustic solution in the piping.

The temperature to cause caustic SCC in carbon steel is 50°C–80°C or more, according to caustic concentration (refer to the API RP571). Accordingly, the welded part of carbon steel that is exposed to this temperature range must be executed post-weld heat treatment (PWHT, stress-relieving) to avoid SCC. Through a hardness check of the sample, stress-relieving on some parts seemed improperly performed. By the above mentioned analysis results, the inspection team reported that the frequent leaks happened due to caustic SCC.

Solution. To fundamentally prevent caustic SCC, a material upgrade to "Ni alloy" could be considered. Alternatively, precise temperature control of the caustic line through electric tracing rather than steam tracing could be selected. Both improvements are costly. The technical staff checked the freezing point of the 25 wt% caustic (**FIG. 13**) and compared it (–20°C) with the regional lowest temperature in the city (–8.5°C). It was not possible that the caustic solution would reach freezing point in that area. Finally, they eliminated stream tracing from the caustic makeup line and have not experienced this type of piping damage since.

Lessons learned. By carefully reviewing process conditions and chemical properties, optimum solutions to corrosion issues can be found. Alternative countermeasures can be considered, but engineers should also consider and compare their costs and effectiveness.

Takeaways. This article has highlighted two key lessons for engineers:

- Close cooperation between corrosion engineers and process and design engineers is essential to determine the best and most economical solution for a corrosion and material issue.[3] The process engineer should provide exact process conditions and operating concepts, including normal operations and abnormal cases (shutdown, startup, turndown and emergency), for each area of concern. This is a starting point for trouble-shooting.
- All discipline engineers (process, mechanical, piping, inspection, materials, etc.) should ensure consistency of design documents at each step, such as basic design by license companies, front-end engineering and detailed engineering by engineering companies, and vendor design. Careful review is required on material, operating and design conditions, stream compositions (including corrosives), piping spec breaks, etc. Reference document checks, such as P&IDs and equipment drawings from other companies, are helpful if no similar process or equipment is in use in existing plants. Equipment inspection must be performed by owner-side engineers at the construction stage; they are responsible for a new process or equipment as an asset owner. HP

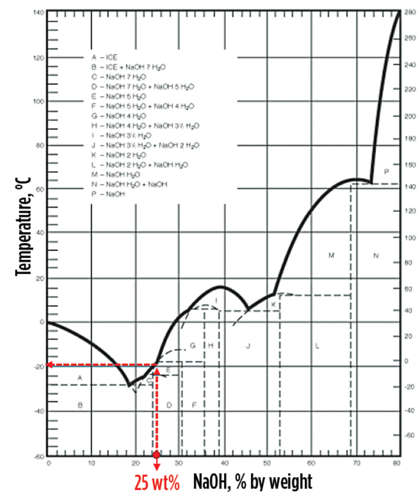


FIG. 13. Freezing point curve for caustic soda (NaOH) solutions.

LITERATURE CITED

1 American Iron & Steel Institute, "The role of stainless steels in petroleum refining," April 1977.
2 UOP, "Corrosion and fouling management course," Des Plaines, Illinois, June 1995.
3 Fontana, M., *Corrosion engineering,* 3rd Ed., McGraw-Hill Companies, 1986.

GI WON PARK works in the upstream division of Kuwait Oil Co. in Kuwait. He has 26 yr of experience in process design and operation and holds two U.S. patents and one international patent in process technologies. He has worked on various projects and revamps, including three billion-dollar projects for grassroots refining (SK Energy) and petrochemical plants as a Senior Process Engineer and Engineering Team Leader. As a corrosion engineer, he is also involved in corrosion failure investigations and reliability improvement task forces. Mr. Park holds a BS degree in chemical engineering.

LEE NICHOLS, Editor-in-Chief/Associate Publisher

The 1970s: Crises, clean air, plastic bottles and the DCS

The 1970s were marked by several historical events that affected not only the hydrocarbon processing industry (HPI) but nations around the world. The decade witnessed two oil crises that would disrupt the global supply of oil and increase prices substantially. New regulations by the U.S. and Europe ushered in an era of clean fuels standards that are still in transition today. Novel technologies introduced in the 1970s revolutionized emissions reduction from vehicles' tailpipes, advanced process controls and automation, and changed the way society drinks carbonated beverages.

The decade also witnessed advances in catalytic processing technologies, such as the commercialization of catalytic dewaxing, wax hydroisomerization and continuous catalytic reforming (CCR).[135] For example, Mobil developed the first catalytic dewaxing process in the mid-1970s. The technology—referred to as the MDDW process—utilized the company's Zeolite Socony Mobil-5 (ZSM-5) catalyst to increase the cold flow properties of diesel (the invention of the ZSM-5 catalyst was detailed in the History of the HPI section in the May issue).[136] In 1971, UOP began operations on the first CCR Platforming unit at the Coastal States refinery in Corpus Christi, Texas (U.S.). According to literature, the CCR section enabled refiners to continuously remove coke accumulating on the catalyst. This allowed lower reforming reaction pressures to increase reformate and hydrogen yields, higher reaction temperatures to achieve higher octane levels for gasoline blending—thus enabling lead-free gasoline—and increased production of aromatics for use as petrochemical feedstocks.[137]

Two crises stress the importance for energy security. The 1970s were rocked by two global crises: The oil embargo of 1973 and the oil crisis of 1979. These two events had detrimental effects on oil importing nations around the world, as well as stressed the importance of energy security.

Oil embargo of 1973. The first oil crisis to affect the global economy in the 1970s was the Organization of Petroleum Exporting Countries' (OPEC's) oil embargo in 1973–1974—the creation of OPEC was detailed in the History of the HPI section in the May issue of *Hydrocarbon Processing.* The embargo was a retaliation against countries that supported Israel during the Yom Kippur War with Syria (i.e., the U.S., Canada, Japan, and a few African and Western European nations).[138] It banned petroleum exports to targeted countries and incorporated crude oil production cuts, leading to a quadrupling of oil prices—oil prices increased from \$3/bbl to nearly \$12/bbl by early 1974, an increase of 300%.[139]

The embargo had detrimental effects on nations that were dependent on foreign oil to satisfy domestic demand. Many nations enacted oil rationing programs, as well as banning fuels usage (e.g., driving, flying) on various days. The high price of oil even led several countries to the brink of recession and proved that oil could be used as an economic weapon.[139]

In March 1974, peace talks between Israel and Syria led to the eventual lifting of the oil embargo. However, this would not be the last time within this decade that the world would be caught in a global oil price crisis.

Oil crisis of 1979. The second crisis that significantly affected global oil prices in the 1970s was due to the Iranian Revolution. The revolution, which began in early 1978 and ended a year later, led to the toppling of the country's leader, Shah Mohammed Reza Pahlavi, and installed Sheikh Khomeini as grand ayatollah. The year-long revolution was responsible for knocking approximately 4.8 MMbpd of oil production offline. Although this represented only 7% of the world's oil production at the time, it led to global oil prices nearly doubling to \$39/bbl.[140]

In a little more than 6 yr removed from the oil embargo crisis, history began to repeat itself. Several countries rationed supplies, governments invested billions of dollars in research to find an alternative to oil, and many countries either switched or began to explore switching domestic power generation from oil to other feedstocks, such as coal, natural gas or nuclear.[140]

Both the 1973 oil embargo and 1979 oil crisis had dramatic effects on the global marketplace. However, the underlying theme of these global events brought to light the necessity for energy security, a concept that continues today.

The Clean Air Act of 1970 ushers in a new era of environmental awareness/action. As countries modernized and produced fuels and products for domestic and international markets, a lingering challenge could be felt by local populations: air pollution. Not only did manufacturing plants, refineries, chemical and petrochemicals facilities, factories and other industrial operations produce air pollutants, gasoline burned in internal combustion engines (i.e., automobiles) filled the skies with pollutants such as hydrocarbons, carbon monoxide and nitrogen oxides.

In efforts to reduce air pollution, the U.S. Environmental Protection Agency (EPA) initiated a series of laws and amendments to various industries. The first federal legislation passed to address air pollution in the U.S. was the Air Pollution

Control Act of 1955. Although the law did not tackle air pollution directly, it provided funding for research relating to air pollution control.[141] Eight years later, the U.S. EPA passed the Clean Air Act of 1963, which enabled the U.S. government to take direct action to control air pollution.[141] In 1965, an amendment to the Clean Air Act of 1963—the Motor Vehicle Air Pollution Control Act—created the first federal set of standards for vehicle emissions.[142]

The Clean Air Amendments of 1970 significantly strengthened federal authority to regulate emissions from both industrial and mobile sources (**FIG. 1**). This amendment included the following major components[141]:

- It established the National Ambient Air Quality Standards (NAAQS) for pollutants in outdoor air that can be harmful to the public or the environment—i.e., carbon monoxide, lead, particulate matter, ozone, nitrogen dioxide and sulfur dioxide.[143]
- It established New Source Performance Standards to determine how much air pollution should be allowed by different industries.
- It established the National Emission Standards for Hazardous Air Pollutants to cover all air pollutants not covered by the NAAQS.
- It called for aggressive air pollution reduction goals—some as high as 90%—for the mobility sector.

The significance of the Clean Air Act of 1970 was that it gave the U.S. EPA enforcement authority over domestic emissions levels, as well as required U.S. states to issue plans (State Implementation Plans) on adhering to national air pollution standards. This model is still in use today. The Clean Air Act had several additional amendments added to it over the next 30 yr, including major additions during the 1990s to address acid rain, ozone depletion and toxic air pollution, as well as establishing Reid vapor pressure standards and new regulations on fuels sold during the months of May–September (i.e., summer-grade fuel).[141]

Although the Clean Air Act was intended to reduce air pollution, especially from the automobile industry, challenges existed on how to mitigate pollutants from an automobile's tailpipe. A solution was put forth in the mid-1950s but did not fully materialize for the auto industry until the mid-1970s. This technology can still be found on nearly every vehicle in use today: the catalytic converter.

The catalytic converter. Although prototypes of catalytic converters were introduced in France in the late 1800s, the modern catalytic converter was first patented in the mid-1950s by a well-known pioneer in the refining industry, Eugene Houdry. Houdry's pioneering work in the creation of catalytic cracking was detailed in the History of the HPI segment in the February issue of *Hydrocarbon Processing*.

FIG. 1. U.S. President Richard Nixon signs the Clean Air Act of 1970, which called for a significant reduction in air pollutants from industrial and mobility sectors. Photo courtesy of the U.S. National Archives.

Houdry began research and development on this technology after studies were released that showed alarming increases in smog in the Los Angeles, California (U.S.) area. These Los Angeles area smog studies in the early 1950s also played a part in similar studies in Western Europe. Around 1956, both French and German scientists were engaged in research to mitigate smog in several major cities in France and Germany.[144] These scientists noticed that several of their respective urban areas suffered from dense air pollution similar to that referenced in the Los Angeles smog reports. Both teams' research into mitigating vehicle emissions eventually led to the implementation of Directive 70/220/EEC in 1970.[145] This ground-breaking piece of legislation was the impetus to setting emissions standards for light- and heavy-duty vehicles in Europe. The directive eventually led to the introduction of the Euro 1 standard in 1992 (implemented for passenger cars in 1993), the removal of leaded petrol from filling stations in Europe and the adoption of three-way catalytic converters.[144,146,147] European emissions standards (i.e., Euro 1–6 and Euro I–IV; Euro 7/VII are expected to be implemented in the mid-2020s)[145,148] would become a global standard for many countries around the world over the next few decades in efforts to adhere to clean fuels regulations.

In the U.S., Houdry was concerned that emissions from smokestacks and automobile exhaust were leading to significant air pollution.[149] To reduce emissions from these sources, Houdry created the company Oxy-Catalyst to develop catalytic converters. His first designs were aimed at mitigating emissions from smokestacks. This effort was followed by the development of catalytic converters for low-grade gasoline-powered forklifts used in warehouses.[150] In the mid-1950s, Houdry fixed his sights on developing catalytic converters for automobile engines. His technology was patented under the title Catalytic Apparatus to Render Non-Poisonous Exhaust Gases from Internal Combustion Engines on April 17, 1956 (**FIG. 2**).[151]

However, the widespread adoption of catalytic converters by the automobile industry did not take effect until the passing of the U.S. Clean Air Act and subsequent amendments. These laws dictated strict regulations on vehicle emissions, as well as the continued removal of lead from gaso-

line—incorporating tetraethyllead (TEL) into gasoline was first used in the 1920s to prevent knocking in internal combustion engines. The first TEL reduction standards—part of the U.S. NAAQS standards—were passed into law in the early 1970s. The recognized adverse impacts of emissions from leaded gasoline on human health would lead to the eventual removal of lead from gasoline over the next few decades—the U.S. banned leaded gasoline in on-road vehicles in 1996.[152] Lead was also detrimental to the operation of catalytic converters. Lead acts as a catalyst poison by forming a coating on the catalysts inside the converter, leading to inactivity and increased emissions.[149] Numerous countries in Asia, Africa, Europe and South America followed suit, and, in July 2021, the last batch of leaded gasoline was sold in Algeria. This occasion marked the end of the use of leaded gasoline globally.[153]

After the adoption of the Clean Air Act, automobile manufacturers began producing new lines of vehicles that included catalytic converters. However, the Clean Air Act amendments of the 1970s put stringent restrictions on the removal of carbon monoxide, hydrocarbon and nitrogen oxide emissions. Catalytic converters available at the time were able to reduce carbon monoxide and hydrocarbon emissions but not nitrogen oxide. This challenge was solved by a group of engineers working at Engelhard Corp. (now part of BASF) in Iselin, New Jersey (U.S.). This group was led by chemists Carl Keith and John Mooney. Their revolutionary three-way catalytic converter—introduced in 1973—was able to reduce all three pollutants from a vehicle's tailpipe. According to literature, the technology used rare-earth and base metal oxide components in the catalyst together with platinum and rhodium in a ceramic honeycomb, with tiny passages coated with the catalytic material.[154] This enabled their design to remove carbon monoxide, hydrocarbon and nitrogen oxide in a single catalytic component.[154] The three-way catalytic converter was installed in most vehicles in 1976 and is still in use today.

The evolution of the distributed control system. In 1959, Texaco started operations on the first digital control computer at a refinery. This system—a Thompsom Ramo Wooldridge RW-300 computer—became the first fully automatic, computer-controlled industrial process and ushered in the computer-integrated manufacturing era in the HPI. A detailed account of this technology was published in the History of the HPI section in the April issue of *Hydrocarbon Processing.*

Additional technologies, such as programmable logic controllers (PLCs), were incorporated into plant operations in the late 1960s/early 1970s. These devices were pioneered by Richard (Dick) Morley of Bedford Associates (now part of Schneider Electric) and Odo Josef Struger of Allen-Bradley (now part of Rockwell Automation). Both inventors are known as the fathers of PLCs—Struger even coined the acronym PLC.[155] A history of the PLC is detailed in the History of the HPI section in the May issue of *Hydrocarbon Processing.* Allen-Bradley also introduced Data Highway in 1979, which was the first plant-floor network designed to support remote programming and messaging between computers and controllers, replacing miles of wiring in plant operations.[156]

In 1975, another revolutionary technology was unveiled to optimize refining and petrochemical plant operations, the distributed control system (DCS). The first DCSs were introduced by Honeywell and Yokogawa. Bristol (now part of Emerson Process Management) also introduced the UCS3000 in 1975, which was the first microprocessor-based universal controller.[157] Prior to the DCS, plant operations were controlled via board operation (i.e., controllers were mounted on large instrument panels). However, through the evolution and wide-scale availability of microcomputers and microprocessors, the DCS was created to control manufacturing processes in several industries, including oil refining and petrochemicals production.[157]

Honeywell and Yokogawa both introduced their own DCSs around the same time—Yokogawa created CENTUM (**FIG. 3**), while Honeywell introduced the TDC2000 platform. According to literature, Yokogawa's journey to the DCS included applying microprocessors to control systems. These control systems were divided into three basic components: human-machine interface, controllers and control bus. The system was named DCS and was instrumental in controlling various functions of plant operations (e.g., flow).[158]

In the early- to mid-1970s, Honeywell worked extensively at optimizing automation technologies, as well as focusing on advancing process controls. The company introduced the TDC2000 (TDC stood for total distributed control) system in 1975. This system provided a centralized view of processes within the plant and utilized a data highway that could link various sensor data to a central location.[159] Plant personnel could monitor and modify several control loops in a single system. TDC2000 was used globally for a decade until being replaced by TDC3000 in 1985, followed by Experion in the 2000s.

In 1978, Valmet introduced the Damatic Classic automation system, which was installed at Pankaboard's board mill in Lieksa, Finland. The DCS operated for nearly 40 yr at that location before being replaced by the latest iteration (Valmet DNA) in 1998.[160]

Other digital companies introduced

FIG. 2. Eugene Houdry holding a small catalytic converter. Photo courtesy of Sunoco and the Science History Institute.

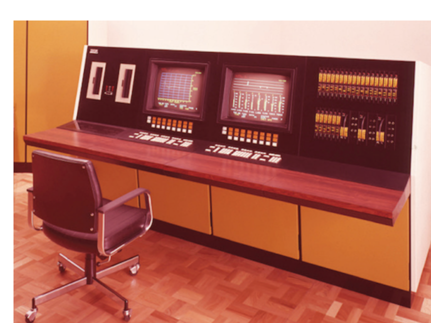


FIG. 3. Yokogawa introduced the CENTUM DCS in 1975. Photo courtesy of Yokogawa.

new technologies during the 1970s and 1980s to optimize process controls and automation for the HPI. In the late 1970s, the Massachusetts Institute of Technology (MIT) created an Energy Laboratory to facilitate collaboration between university and industry.[161] This effort materialized out of the energy crisis of the 1970s. Led by MIT Professor of Chemical Engineering Larry Evans and funded by the U.S. Department of Energy, the Advanced System for Process Engineering (ASPEN) project began in 1977.

According to literature, the ASPEN project set about to develop a general-purpose simulation system that could be used by chemical engineers across the entire process industries. The result of the project was the next-generation process simulator named ASPEN. This technology could simulate large, complex processes involving highly non-ideal chemical components, coals and synthetic fuels, as well as electrolyte and multiphase systems.[161]

In 1981, the software was commercialized by the new company, AspenTech, which released its first product, Aspen Plus, in 1982.

Several direct digital control technologies were released in the 1970s, which included Foxboro's (now part of Schneider Electric) FOX 1 system for plant monitoring and process control, Fisher Controls' (now part of Emerson) DC² system and PROVOX DCS, Taylor Instrument Co.'s and Baily Controls' (both companies are now part of ABB) 1010 system and 1055 system, respectively.[162,163]

Process automation continued to evolve over the next several decades, including the move to ethernet-based networks, fieldbus installations, wireless systems and protocols, increased cyber defenses, remote transmission, and many other advances to optimize plant operations.

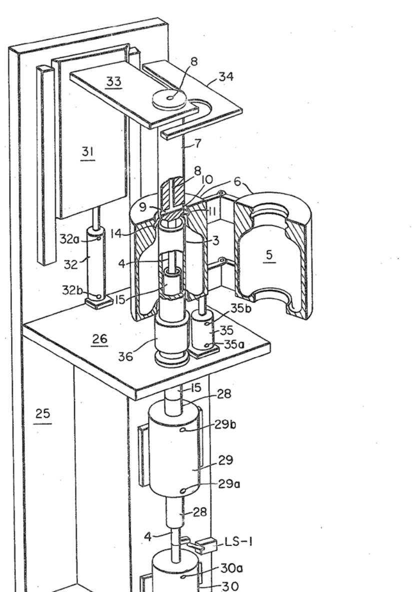


FIG. 4. A perspective view of Wyeth's invention as submitted in his patent. Photo courtesy of the U.S. Patent Office.[170]

Polyethylene terephthalate: Solving the carbonated liquids container challenge. In 1941, DuPont scientists John Whinfield and James Dickson expanded on Wallace Carothers'—a fellow DuPont colleague—work on synthetic fibers. Carothers' research was instrumental in the discovery of neoprene, nylon and other synthetic fibers. These discoveries were detailed in the February issue's History of the HPI section of *Hydrocarbon Processing.*

Through their research, they discovered how to condense terephthalic acid and ethylene glycol into a new polymer that could be drawn into a fiber.[164] Their work eventually led to the development of polyethylene terephthalate (PET). Whinfield and Dickson patented their discovery in Great Britain in mid-1941 (and later in the U.S. in 1945);[165] however, due to wartime secrecy, the invention was not made public until several years later.[166] PET would become the basis for many products used in everyday life, and, today, PET is the fourth most produced polymer. One of the primary reasons for its popularity is its stretchability into long hard fibers, which makes it ideal to produce films and containers, among other items, that are lightweight, hard and durable. Using blow molding on PET created a product in the early 1970s that would revolutionize how societies enjoy different beverages: the plastic bottle.

The first plastic bottle was created in the late 1940s by cosmetic chemist Jules Montenier. At the time, Montenier was trying to find a suitable container for his liquid antiperspirant called Stopette—prior to his invention, antiperspirants were applied as a cream or in liquid form by dabbing it on using an applicator or pad.[167] He turned to a new chemical polymer discovered approximately a decade before called polyethylene (PE)—a detailed account of the discovery of PE was published in the History of the HPI section in the February issue. In 1947, Montenier partnered with the Plax Corp. of Hartford, Connecticut (U.S.)—the company used blow molding to manufacture plastic Christmas tree ornaments.[167] Their partnership produced the Stopette spray bottle, which was first commercially sold in July 1947.[167] This event marked the beginning of plastic containers competing against glass.

However, plastic containers remained expensive until the invention of high-density PE (HDPE)[168] in the 1950s by J. Paul Hogan and Robert L. Banks while working at Phillips Petroleum Co. in Bartlesville, Oklahoma (U.S.)—the discovery of HDPE is detailed in the History of the HPI section of the April issue. Several new uses of plastic bottles were commercialized over the next two decades, including the plastic milk bottle (patented by Roy Josephsen, Joseph Tino and Charles Fulcher of W. R. Grace & Co.) in 1965.

Like Whinfield and Dickson, Nathanial Wyeth also worked at DuPont. Prior to the late 1960s, he invented several products for the company, including a machine that built dynamite cartridges automatically, which kept workers from inhaling poisonous nitroglycerin powder; and a machine to manufacture Typar, a polypropylene (PP) fabric used in industrial sectors such as construction.[169]

In 1967, Wyeth began experimenting with the possibility of using plastic bottles to store carbonated beverages. Conventional wisdom at the time was that plastic bottles could not hold the pressure of carbonated beverages and would explode. To test this theory, Wyeth filled a plastic detergent bottle with ginger ale, sealed it and placed it in his refrigerator. According to literature, the next morning, the bottle had swelled so much that it was lodged between the refrigerator shelves.[169] This experiment proved to Wyeth that a stronger plastic was needed to withstand the pressure of carbonated liquids.

His initial work was with PP; however, he switched to PET due to its superior elastic properties.[169] After several experiments, Wyeth invented a machine that produced a "hollow, biaxially-oriented, thermoplastic."[170] This machine would

strengthen the plastic by creating a mold that had nylon thread running in a diamond crisscross pattern. When the mold was pressed, the molecules aligned in a biaxial fashion.[169] This created a light, clear and resilient product that could withstand the pressure of carbonated liquids. On May 15, 1973, Wyeth received a U.S. patent for his biaxially-oriented PET bottle machine (**FIG. 4**).

Although PET plastic bottles were more expensive than glass when first introduced into the market, they had many more benefits, such as they were lighter, they were not easily breakable and they could be resealed. Eventually, due to increased manufacturing, the cost for PET plastic bottles decreased significantly.[171] Companies like Coca-Cola and Pepsi brought PET plastic bottles to the global masses, and PET plastic bottle usage has soared globally over the past several decades. In 2021, more than 580 B PET plastic bottles were produced (an increase of nearly 100 B/yr since 2016), reaching a total market value of nearly $40 B—industry reports forecast the PET plastic bottle market reaching more than $50 B by 2027.[172,173]

Infrastructure rises from the Saudi desert: Jubail, Yanbu and the master gas system. In 1975, Saudi Arabia's government commissioned the construction of two new industrial cities, one on each of its coasts—Jubail in the east and Yanbu in the west. These cities were the results of the country's growing wealth from oil production and global trade and would serve as major industrial complexes to produce refined fuels and petrochemical products to satisfy domestic demand and for export.

Around the same timeframe, Aramco—the company would not adopt the name Saudi Aramco until the late 1980s—began work on the country's master gas system.[174] The system's goal was to gather and utilize associated natural gas that was being flared (wasted) from domestic production and use it as a low-cost fuel for industrialization.[175] This capital-intensive project included the construction and operation of gas gathering infrastructure, treating and processing facilities, and a transmitting system. By the mid-1980s, the master gas system was able to produce up to 2 Bft^{3}d of natural gas.[175] Over the next 40 yr, the company significantly expanded the system's total capacity, with the ability to produce approximately 12.5 Bft^{3}d of natural gas by the early 2020s.

Jubail Industrial City. Jubail's origins date back more than 7,000 yr and garnered fame in 1933 as the initial landing spot for Standard Oil of California (now Chevron) geologists in their search for oil in the country.[176]

In the mid-1970s, Jubail was little more than a fishing village; however, it had several benefits for the country. The city's location was ideal for shipping, it had ample water supplies to cool industrial plants and it was near crucial domestic oil production fields.[177]

The scope of the megaproject was to convert Jubail into a large-scale industrial city. The Saudi government selected two agencies to oversee the city's construction: The General Petroleum and Mineral Organization (PETROMIN) and the Saudi Basic Industries Corp. (SABIC). The project developers selected American-based engineering, construction and project management firm Bechtel to design and build the industrial city. Jubail Industrial City was an effort by the Saudi government to reach self-sufficiency in refined and petrochemical products.

The city, which covers more than 1,000 km^2, includes a multitude of industrial infrastructure, including the 440,000-bpd SATORP refinery (a JV between Saudi Aramco and TotalEnergies) and the SADARA petrochemical complex (a JV between Saudi Aramco and the Dow Chemical Co.).

Yanbu. On the country's west coast, the Saudi government decreed the construction of a second industrial city in Yanbu. The city's origins date back more than 2,500 yr when it was used as a staging point on the spice and incense route from Yemen to Egypt and various countries around the Mediterranean.[178] This sister industrial city to Jubail would be smaller, but due to its proximity on the Red Sea, would be crucial as an import/export port for the country. Over the next several decades, additional hydrocarbon processing facilities would be built, including refineries, petrochemical plants and other supporting infrastructure (e.g., pipelines, storage).

Today, Jubail and Yanbu are the first- and fourth-largest industrial cities, respectively, in the world.[179] HP

LITERATURE CITED

135 Machinery Lubrication, "The advent of modern hydroprocessing: The evolution of base oil technology—Part 2," online: https://www.machinerylubrication.com/Read/493/base-oil-technology

136 Hilbert, T., M. Kalyanaraman, B. Novak, J. Gatt, B. Gooding and S. McCarthy, "Maximizing premium distillate by catalytic dewaxing," *Digital Refining*, February 2011.

137 Honeywell UOP, "50 years of CCR Platforming," online: https://uop.honeywell.com/en/products-and-services/catalysts/refining-catalyst/ccr-platforming

138 U.S. Department of State, "Oil embargo, 1973–1974," Office of the Historian, online: https://history.state.gov/milestones/1969-1976/oil-embargo#:~:text=The%20onset%20of%20the%20embargo,stability%20of%20whole%20national%20economies

139 Wikipedia, "1973 oil crisis," online: https://en.wikipedia.org/wiki/1973_oil_crisis#Consequences

140 Federal Reserve History, "Oil shock of 1978–79," November 22, 2013, online: https://www.federalreservehistory.org/essays/oil-shock-of-1978-79

141 Wikipedia, "Clean Air Act," online: https://en.wikipedia.org/wiki/Clean_Air_Act_(United_States)#History

142 USLegal, "Clean Air Act," online: https://environmentallaw.uslegal.com/federal-laws/clean-air-act/

143 U.S. EPA, "Reviewing NAAQS: Scientific and technical information," online: https://www.epa.gov/naaqs

144 Smith, A. and H. Davies, "A review of the history of emission legislation, urban and national transport trends and their impact on transport emissions," *Transactions on the Built Environment*, vol. 23, 1996, online: https://www.witpress.com/Secure/elibrary/papers/UT96/UT96028FU.pdf

145 Wikipedia, "European emissions standards," online: https://en.wikipedia.org/wiki/European_emission_standards

146 The Council of European Communities, "Council Directive 70/20/EEC," Office Journal of the European Communities, March 20, 1970, online: https://eur-lex.europa.eu/legal-content/EN/TXT/PDF/?uri=CELEX:31970L0220&from=EN

147 TransportPolicy.net, "EU: Light duty emissions, online: https://www.transportpolicy.net/standard/eu-light-duty-emissions/

148 European Automobile Manufacturers' Association, "ACEA proposals for Euro 7 and Euro VII emission standards," January 28, 2022, online: https://www.acea.auto/publication/acea-proposals-for-euro-7-and-euro-vii-emission-standards/

149 CatalyticConverters.com, "History of the catalytic converter," online: https://www.catalyticconverters.com/history-2/history/

150 Wikipedia, "Catalytic converter," online: https://en.wikipedia.org/wiki/Catalytic_converter#History

151 Houdry, E., "Catalytic structure and composition," U.S. Patent No. 2,742,437, U.S. Patent Office, April 17, 1956, online: https://upload.wikimedia.org/wikipedia/commons/6/6c/US2742437_Houdry_Auto_Catalyst.pdf

152 U.S. EPA, "EPA takes final step in phaseout of leaded gasoline," January 29, 1996, online: https://archive.epa.gov/epa/aboutepa/epa-takes-final-step-phaseout-leaded-gasoline.html

Complete literature cited available online at www.HydrocarbonProcessing.com

BASF Refinery Catalysts: Innovating the future from in-situ zeolites to renewable feedstocks

Global demand for energy today is growing rapidly due to an increasing population and urbanization, combined with a rising middle class. Sources of energy supply have evolved throughout human history, and to meet society's needs the global energy supply now needs greater diversification and sustainability. Since the 1865 founding of BASF, the company has been involved with organic and inorganic chemistry and the creation of innovative products for the petrochemical and the energy industry.

BASF has supported the refining industry with FCC catalysts and additives since 1972, a year after launching its gas treatment business (OASE®). BASF also provided new innovations to refiners which included fuels additives, guard beds for sulfur, chloride and fluoride removal, SHU catalysts (selective hydrogenation catalyst), Claus catalysts, dearomatization catalysts, and deolefination catalysts.

The fluid catalytic cracking (FCC) unit was a critical innovation to meet the modern world's energy demand. Donald Campbell, Eger Murphree, Homer Martin, and Charles Tyson are credited with the landmark invention of FCC, which revolutionized the refining industry by providing an efficient process to increase the yield of high-octane gasoline from crude oil. New catalysts were needed for the FCC unit to make the process work.

BASF extended upon our Thermofor catalytic cracking (TCC) catalyst (also known as Moving Bed Catalyst) to create a novel FCC catalyst product using an in-situ zeolite crystallization process. Walter Haden and Frank Dzierzanowski invented BASF's unique in-situ-zeolite growth technology to produce BASF's FCC catalysts. Their unique catalyst invention was awarded a US patent in 1968. In 1972, BASF rapidly launched a new production facility to introduce this novel technology to the refining industry. BASF's FCC technology enabled the industry to capitalize on the emerging design trends and new technology designs installed in turnarounds and new construction in the late 1970s and until today. Design features such as short contact time riser reactors, new nozzle atomization designs and close coupled cyclone technology were introduced to the market, which enabled BASF's novel high activity FCC catalysts to deliver the performance required in the market.

The emerging need for increased selectivity to propylene and butylene also have been supported with new technologies, such as the 2019 introduction of the multiple framework topology (MFT) technology, which resulted in Fourte®, Fourtune® and Fourtitude® product introductions for butylene selectivity in VGO and resid feedstocks. Improved Zeolite Y (IZY) technology enhanced our Distributed Matrix Structure (DMS) technology to deliver leading catalyst activity and improved diffusion capability. Recent improvements in metals passivation and trapping have resulted in the introduction of Valor® for vanadium passivation and our Boron Based Technology (BBT) for nickel passivation. BoroCat®, BoroFlex® and Borotec® FCC catalysts are all part of the BBT family of product delivering the novel innovation for use in resid applications.

Additionally, BASF has continued to deliver innovative additives to meet the performance and environmental needs of the FCC unit. EnviroSOx® supports SOX emissions reductions while our ZIP® and ZEAL® products are designed to enhance propylene yields. Enable® is our newest CO promoter which also has a lower NOx generation profile than typical platinum CO promoters.

The transition to sustainable energy requires continued innovation from BASF and others to support the FCC unit changes as refining remains a key part of the global energy portfolio. BASF's research and development expertise continues to lead the way developing new innovative and creative catalysts and process solutions to enable the use of new alternative feeds – such as biomass, bio-oils/ fats, and pyrolysis oils. The emergence of regulations supporting decarbonatization have now created an opportunity in refining to use these more sustainable feedstocks and eventually the potential to evaluate new carbon capture technologies such as OASE® for stack gas carbon capture. The increasing use of renewable energy as a part of the refining complex also allows refiners to consider production of green or blue hydrogen to support their hydrogen requirements while significantly reducing their carbon footprint.

Meeting the energy needs of the future requires a diverse network of refiners, licensors, experts, and suppliers to work together to meet the sustainability and energy challenges for our future. At BASF, we create chemistry for a sustainable future.

CONTACT INFORMATION

Address: BASF Corporation,
33 Wood Avenue,
08830-0770 ISELIN, United States
Email: mark.schmalfeld@basf.com
Website: www.catalysts.basf.com/refinery-catalysts

LEE NICHOLS, Editor-in-Chief/Associate Publisher

Excerpts from the 1970s: Maintenance, optimization, safety and digital technologies

You can predict turbomachinery reliability
J. S. Sohre, January 1970

Reliability factor curves have been drawn as a guide to probable trouble in components or entire turbomachinery systems. Designers can use reliability factors to upgrade weak system components. Maintenance can use the factors to pinpoint probable trouble areas.

Low-sulfur fuels are different
C. W. Siegmund, February 1970

A change to low-sulfur fuel oils can change the requirements for storing and pumping these fuels. Fuels from low-sulfur crudes will have a higher viscosity, while desulfurization of high-sulfur stocks will produce lower viscosity but higher costs.

What's ahead for aromatics?
S. Field, May 1970

In this decade (1970s), chemicals will require 230,000 bpd and gasoline 1 MMbpd of aromatics. Refiners will face new problems in butene, xylene and toluene production, conversion, processing and extraction to meet this demand.

Refiners face challenges of 1970s
J. D. Wall, December 1970

Challenges facing the refining industry will cause shifts in locations of refining centers, product distributions and processing styles. Changes will take place to accommodate growing demand for more consumer products and chemical feeds. The needs are clear. The solutions are less clear.

A bigger role for cat reforming
H. L. Hoffman, February 1971

Catalytic reforming of naphthas to produce aromatics is growing at an accelerated pace. The HPI looks to this process for aromatics to improve antiknock ratings of motor fuels and to furnish feeds for petrochemicals.

Optimize olefin cracking coils
J. L. de Blieck and A. G. Goossens, March 1971

Based on simple kinetics and extensive pilot and commercial plant operations, the radiant cracking coils can be optimized for product pattern and investment.

Natural gas grows in world use
J. Salkeld, April 1971

The U.S. and the USSR have long dominated the use of natural gas. Now, with better pipelines and LNG, gas is on the move. It is becoming a bigger factor in the energy picture of most world areas. The article examines what can be expected in the judgment of an authoritative market analyst.

Synthetic fiber feedstocks
P. Leprince, July 1971

Synthetic fibers—nylons, polyesters, acrylics and polyolefins—show growth rates above 10%/yr. With this fiber growth and its increasing demand on HPI feedstocks, present data indicates ample raw material supplies with minimum supply-demand dislocations.

How styrene grows in Europe
July 1971

From 500 MMlb of styrene converted into plastics in 1960, markets have grown to about 2.5 Blb today. A future annual growth rate for styrene monomer of 10%–12% is expected.

Flue gas desulfurization technology
A. B. Welty, Jr., October 1971

How do we keep sulfur out of the air? Flue gas desulfurization is one way that many are considering, and a few are trying. This article is an analysis of the possibilities for making this approach work. It is "must reading" if you have a flue gas problem.

Wastewater treatment
M. R. Beychok, December 1971

Stronger laws and tougher enforcement make design of the wastewater treatment system more important, but state-of-the-art processes can meet the challenge when properly used by the design engineer.

Better gasoline chromatography
H. Boer and P. van Arkel, February 1972

Multistage chromatography is used to determine the paraffin-naphthene-aromatic composition of products in the naphtha boiling range.

Optimize the ethylene complex
A. J. Gambro, K. Muenz and M. Abrahams, March 1972

This article details how planning, optimization and process selection can maximize profits for a 1 Blb/yr ethylene petrochemical refinery.

Can olefins meet '70s needs?

R. I. Silsby and N. E. Ockerbloom, March 1972

To fill a potential world demand of 60 MMlb/yr by 1980, olefins plants must meet these prerequisites and challenges: plant size, feedstocks, technology, engineering, inflation and finance.

Make SNG from coal?

G. E. Klingman and R. P. Schaaf, April 1972

Coal is a vast reservoir of energy that cannot be fully tapped until processes are developed for conversion to a useful form, such as substitute natural gas (SNG). The race is on to get the best process first. This article examines the frontrunners.

Regenerate reformers continuously

B. J. Cha, R. Huin, H. Van Landeghem and A. Vidal, May 1973

Continuous regeneration of catalytic reformers offers an economical way to high octanes and high stream factors. New units can be designed to which continuous regeneration is added later.

Which heat recovery system?

A. Mol, July 1973

Check these pros and cons when selecting integrated vs. central waste heat recovery systems for ethylene plants.

How stable is diesel in storage?

M. E. LePera and J. G. Sonnenburg, September 1973

Present specifications may be inadequate to assure suitable diesel fuel storage stability. Laboratory tests are compared with actual 2-yr storage to judge a more suitable specification.

Polybutylene—New isotactic polymer

October 1973

Polymerizing butene-1 using Ziegler-Natta catalyst to crystallize polybutylene makes a unique polymer with interesting and useful physical properties.

How to design a plant firewater system

A. M. Woodard, October 1973

Industries that process large quantities of highly flammable fluids must be protected by well-designed and properly installed firewater systems. This article provides the fundamental design criteria that must be considered. Included are layouts showing a typical process unit fire protection system and a firewater pumping station.

How to improve compressor operation and maintenance

H. M. Davies, January 1974

All centrifugal compressors have their own operating and maintenance problems. However, these machines can be designed and built with operation and ease of maintenance in mind. This article will provide what to consider when selecting a new compressor.

Create by talking with a compute

G. E. Nevill, Jr., R. A. Crowe and J. R. Charles, Jr., March 1974

Tie a digital unit to a terminal, program it with the right words—then converse with it. The result is an aid to innovation in which the computer guides process, and the user implements ideas.

How to apply electric motors in explosive atmospheres

R. L. Nailen, February 1975

Equipment and personnel safety demands consideration of all the facts when an electric motor is to be operated in a hazardous plant environment. This articles provides guidelines that will help avoid many application errors.

LNG goes worldwide

J. D. Wall, April 1975

Continuing planning for new projects is expanding and enlarging the role of LNG in the HPI. Many projects are being considered, most of which are very involved in scope.

Ways to hydroprocess resids

A. G. Bridge, J. W. Scott and E. M. Reed, May 1975

Residuum processes are compared to show hydroprocessing is a commercially vialbe way obtain best liquid yields while meeting modern environmental criteria. Special consideration is given to residual desulfurization and vacuum gasoil/vacuum residual desulfurization for making low-sulfur fuel oils.

Reduce relief system costs

T. W. Whelan and S. J. Thomson, August 1975

The best way to develop an economical relief system design combines a computer program for engineering calculations and applied judgment on realistic relieving quantities.

Startup of a sour gas plant

D. L. Talley, April 1976

Everyone has trouble when they startup a new plant. How they handle the trouble is interesting and educational for all who must do the same thing. This article details Exxon's results in starting up a new gas recovery treating and sulfur recovery system.

From refining to coal gasification: New steels for fuel processing

R. A. Lula, June 1976

New ferritic stainless steels and austenitic steels with higher alloy content have promising corrosion resistant properties for oil refining and coal processing.

Recover energy with exchangers

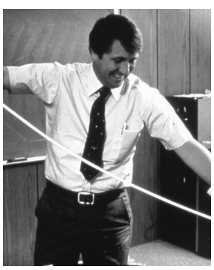
J. B. Fleming, H. E. Duckham and J. R. Styslinger, July 1976

There are useful ways to curtail the consumption of excessive energy in heat exchanger operation. This article details an engineer/constructor's view of the design and application of these essential pieces of equipment and how operating companies can use them to their advantage.

SUMEDHA SHARMA, Technical Editor

Industry Pioneers: Catalysts, polymer processing technology, synthetic lube oils, control systems

ROBERT WALTON GORE



Robert W. Gore was an American engineer, inventor and entrepreneur who is best known for his breakthrough invention of expanded polytetrafluoroethylene (ePTFE). Gore's discovery that PTFE could be transformed into an entirely different physical state led to a phenomenally new direction in material science, resulting in commercially well-known products such as GORE-TEX fabric, a water-resistant and breathable fabric known for its applications in sporting and outdoor activities. Several important products have grown from ePTFE such as new electrical cables, industrial filters, medical implants, textiles woven from ePTFE fiber for space exploration, laminated fabrics for outdoor activities, emergency response, defense and ELIXIR guitar strings.

Gore's father was a Dupont employee and he often experimented with DuPont materials in his basement exploring new ways to use them. While he was a sophomore at the University of Delaware in 1957,[1] Gore helped his father develop a successful process to use PTFE to insulate multiple copper conductors to create the ribbon cable, a product highly applied in the growing computer industry. Gore's process resulted in the product MULTI-TET cable and led his family to found W.L Gore & Associates in 1958,[1] operating from the basement of their home. The company expanded its capacity with the growing demand and applications of MULTI-TET and TETRA-ETCH, a pipe thread tape, and Gore earned his first patent as the inventor. After completing his doctorate in chemical engineering, he joined W.L. Gore & Associates as the technical and research leader. In 1969, while researching a process for stretching extruded PTFE into pipe-thread tape, he discovered that the polymer could be expanded. Gore's discovery of ePTFE resulted from a 'frustrated hard yank' after a series of failed experiments to stretch heated rods of PTFE by 10%. This serendipitous discovery was that instead of slow stretching, the application of a sudden accelerating yank stretched the PTFE by 800%, creating a microporous PTFE that was 70% air.

Gore's earned nine patents for his phenomenal work with fluoropolymers and was elected to the National Academy of Engineering in 1995 for his technical achievements.[2] He was also awarded the highest award in the United States designated for an industrial chemist, the Society for Chemical Industry's Perkin Medal in 2005 and the 2003 Winthrop-Sears Medal, from The Chemists' Club and the Chemical Heritage Foundation, now the Science History Institute.[2]

CHARLES PLANCK AND EDWARD ROSINSKI

Charles J. Plank and Edward Rosinski invented a zeolite catalyst for catalytic cracking that revolutionized the petroleum industry by increasing the yield of gasoline by 40%[3] from every barrel of oil run through a catalytic cracker. Thermal cracking, or the application of heat to petroleum, is the process by which the larger molecules "crack" or break down to form simpler molecules like those found in commercially useful products like gasoline. Plank and Rosinski, while researching catalysts for Mobil Oil (now ExxonMobil) in the 1950s,[4] idealized the use of porous clay-like zeolites that bear microscopic channels close to the hydrocarbon length as catalysts for petroleum cracking. In 1961, it was discovered that certain crystalline zeolites could be combined into a binder and converted into a super-efficient cracking catalyst.[3] Zeolites present superior activity and selectivity at low severity, resulting in significantly high gasoline yield. Moreover, the increased yields are obtained without increasing gas or coke formation, the undesired byproducts of cracking. With higher efficiency and fewer process risks than traditional methods during those times, their process marked a major step forward for the petrochemical industry.

In July 1960, Plank and Rosinski's patent "Catalytic Cracking of Hydrocarbons with a Crystalline Zeolite Catalyst Composite" was submitted and was officially patented on July 7, 1964. Mobil named it "Zeolite Y" and used it in commercial processes in 1964.[4] Through the mid-1980s, nearly 35% of U.S. gasoline was being produced via zeolite catalytic cracking.[4]

Today this catalyst, the first containing crystalline zeolite, is extensively used in all cracking units in the U.S. and around the world. Although catalysts had long been used in oil refining, Plank and Rosinski's catalyst made a significant impact on the efficiency of the cracking process and provided a remarkable increase in gasoline yield from crude oil.

Charles J. Plank was born in Calcutta, India and later moved

to Lafayette, Indiana. In 1936, he received a BS degree in mathematics, chemistry and physics from Purdue University.[5] He later earned an MS degree and in 1942, he received his Ph.D. in physical chemistry from Purdue University.[5]

In 1941, Plank joined the research department of Socony-Vacuum Oil Company, the predecessor of Mobil Oil Corporation. He was promoted as a senior scientist in 1970,[4] the highest scientific post, at Mobil's Research and Development Laboratory. Throughout his career as a scientist and technologist, he was awarded 83 U.S. patents and several hundred in other countries.[5]

Edward Rosinski was born in Gloucester County, New Jersey and aspired to be a chemical engineer while still in high school. Upon graduating in 1939, he joined the Vacuum Oil Company as a petroleum engineer. After working with a couple of instrument companies in the interim, he returned to Socony-Vacuum as a lab technician in 1947, resumed his education, and in 1956 received a BS degree in chemical engineering at Drexel.[6]

In 1972, he was promoted to senior research associate, the company's second-highest scientific post. Rosinski was awarded 76 U.S. patents, of which many were in the field of zeolite catalytic technology.[6]

Rosinski and Plank's paper published in the journal *Industrial and Engineering Chemistry* was voted as one of the 12 most important papers published in the journal.[4] In 1979, Plank and Rosinski were inducted as the 30th and 31st members of the National Inventors Hall of Fame for US Patent No. 3,140,249, "Catalytic Cracking of Hydrocarbons with a Crystalline Zeolite Catalyst Composite."[4]

ALBERT AMATUZIO

Albert Amatuzio was a passionate flyer and a visionary who invented the first synthetic motor oil under his company's name Amsoil Inc. while still serving as a squadron commander of the Air National Guard. His invention brought synthetic lubrication to the automotive market and changed both the automotive and lubrication industries forever.

Amatuzio's entrepreneurial bent surfaced at a young age as he devised several small ventures to support his family during the Great Depression, but his passion for flying led him to join the Naval Corps and then the Merchant Marine. In the post-war period, Amatuzio joined the Air Force, earned his wings and after a hiatus due to family reasons, joined the Duluth unit of the Air National Guard. He served as a fighter pilot for 25 yr and then as a squadron commander. He was honored as the country's top pilot, winning the prestigious William Tell Air-to-Air Shootout competition and the Earl T. Rick Competitive Shootout.[7]

As a pilot, he gained knowledge about how jet engines survived on synthetic oil and envisioned that the same could be applied to other vehicles and equipment that people used in their daily lives. He believed that the same performance benefits would prove invaluable to cars, trucks and other combustion engines. Oil quality during those times was poor—with problems of low heat resistance, contribution to hard-start during cold weather and adverse effects on engine life and performance. He reasoned from his experience that only synthetic oils could avoid these adverse effects and improve engine performance.

Amatuzio's ideas seemed radical and unnecessary at that moment. "They all thought I was at altitude too long without oxygen," Amtuzio joked about his skeptics.[7] However, with his unmatched resilience and tenacity, he dismissed the doubters and began his research and development efforts in 1963.[7] By 1966, he formulated the first synthetic motor oil and founded his company Amsoil.[8] In 1972, Amsoil's tagline 'The First in Synthetics®' was launched as AMSOIL 10W-40 Synthetic Motor Oil became the world's synthetic motor oil to meet American Petroleum Institute's requirements.[8]

Amatuzio had changed the course of the entire automotive lubricant industry. His relentless efforts to bring the best choice to consumers led him to make AMSOIL a technological leader and create the AMSOIL dealer network. His product had met a lot of criticism for being unnecessary, disruptive and "fake," but the founding of the dealer network in 1973 conveyed the benefits of synthetic lubes to consumers.

In 1994, he was honored as the pioneer of synthetic lubrication and inducted into the Lubricant's Hall of Fame.[9] He received the Natchman Award from the Independent Lubricant Manufacturers Association. A community man and a great philanthropist, Amatuzio is remembered through the Albert J. Amatuzio Research Center. The center located in Duluth Depot outlines local service history and includes photographs, journals, stories and biographies of veterans from northeastern Minnesota who served this nation from the Civil War through Iraq and Afghanistan.[9]

STEPHANIE L. KWOLEK

Stephanie Louise Kwolek, an inductee to the National Inventors Hall of Fame and National Women's Hall of Fame, created the first family of synthetic fibers of exceptional strength and stiffness. Kwolek spearheaded the discovery, processing and development of high-performance aramid fibers. Kevlar, the best-known member of this class of fibers, is widely used in more than 200 applications, including protective bullet-proof vests, boats, airplanes, mooring ropes and fiber-optic cables and canoes.

Kwolek was born in New Kensington, Pennsylvania and was encouraged by her naturalist father to develop an early love for nature, math and science. She pursued a BS degree in chemistry from the Carnegie Institute of Technology and wanted to make a career in medicine.[10] She joined DuPont as a researcher at the textile fibers laboratory aspiring to save money for medical school. However, her research focusing on creating stronger and stiffer fibers was extremely challenging and interesting and led to her decision to make chemistry a lifetime career.

Kwolek was working on developing high-performance fibers for extreme applications when she discovered that under certain conditions a large number of molecules of rod-like polyamides line up to form liquid crystalline solutions, which can be spun directly into oriented fibers of very high strength and stiffness. With this breakthrough came the development of Kevlar in 1965, the most acclaimed product of her research—a polymer fiber five times stronger than the same weight of steel and her discovery of a new branch of polymer science—liquid crystalline

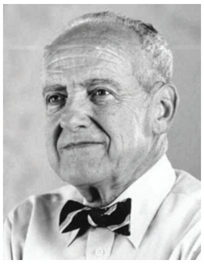
polymers.[10] Her other noteworthy contributions include a low-temperature (0°C–40˚C) condensation process for synthetic fibers.[10] Unlike the conventional melt condensation polymerization process used in preparing nylon, which was typically done at more than 200°C, the new lower-temperature polycondensation processes employed very fast-reacting intermediates, making it possible to prepare polymers that cannot be melted and only begin to decompose at temperatures above 400°C.

Kwolek received over 17 U.S. patents, including one for the spinning process for aramid fibers and five for the prototype from which Kevlar was created in 1965, and won many awards for her invention of Kevlar fiber technology.[11] She was inducted into the National Inventors Hall of Fame in 1994, received the American Innovator Award in 1994, National Medal of Technology in 1996, the Perkin Medal in 1997 and the Lemelson-MIT Lifetime Achievement Award in 1999.[11] In 2003, she was inducted into the National Women's Hall of Fame.[10]

She retired in 1986 but continued to consult for DuPont and served on the committees of the National Research Council and the National Academy of Sciences.[12]

Kwolek continued to mentor women scientists and contributed to science education for young children. One of Kwolek's most cited papers, co-authored with Paul W. Morgan, is "The Nylon Rope Trick" (*Journal of Chemical Education,* April 1959, 36:182–184).[10] It describes a demonstration of condensation polymerization in a beaker at atmospheric pressure and room temperature—which is now a common demonstration in classrooms across the nation. In 2013, Edwin Brit Wyckoff published a childrens' book telling her story as: *The Woman Who Invented the Thread That Stops the Bullets: The Genius of Stephanie Kwolek.*[10]

NATHANIEL WYETH



Nathaniel C. Wyeth was an American engineer and inventor who is credited with the invention of one of the most convenient and readily recyclable plastic products today—the plastic soda bottle. Wyeth invented or co-invented about 25 products and processes in plastics, textile fibers, electronic and mechanical systems.[13]

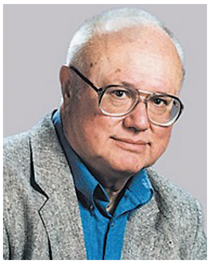
Nat Wyeth was born in Chadds Ford, Pennsylvania, into a family of artists but displayed an early interest in engineering by disassembling clocks and using their parts to make model speedboats, cutting up tin cans and soldering the pieces to make universal joints, and so on. His family recognized the budding inventor's interest and encouraged him. He followed his interests to choose the University of Pennsylvania for its engineering program. During college, Wyeth built a 20-ft-long hydroplane boat that could reach speeds of 50 mph, resting on two pontoons and powered by a Ford V-8 engine.[14] He joined General Motors upon graduation but soon chanced upon an opportunity to work as a field engineer for DuPont Corporation.

During his early days at DuPont, Wyeth excelled by inventing a plug-proof valve for the production machine and was transferred to the mechanical development lab. One of the first machines he designed was for the automatic manufacture of dynamite cartridges, which saved workers from exposure to poisonous nitroglycerin powder. Another notable invention was a machine bearing magnetized rollers, employed in the manufacture of a non-woven polypropylene fabric, Typar®.[13]

By 1967, Wyeth started working on his best-known invention, which began with his curiosity as to why plastic was not used for carbonated beverage bottles.[14] Wyeth was aware that the fabrication process created weak spots in plastic containers, and they were therefore incapable of withstanding carbonation pressure. He took to hands-on experimentation to discover ways to make stronger plastic containers. He knew that stretching out nylon thread strengthened it by forcing its molecules to align. His challenge was to stretch plastic such that its molecules would align in two dimensions, rather than just one—biaxially. He succeeded in doing this by creating a preform mold for the bottle, which resembled a test tube with screw threads running in a diamond criss-cross pattern, instead of single spiral.[13] As the plastic was extruded through this mold, the molecules aligned biaxially—just as Wyeth had intended. The criss-cross flow lines reinforced themselves, creating a uniformly strong product. He also replaced the polypropylene that was used typically for plastic bottles with polyethylene-terephthalate (PET), a polymer with superior elastic properties. He had created a petrochemical product that was light, clear, resilient, safe and eminently recyclable and laid the groundwork for future process developments in preforming, extrusion and manufacture of biaxially oriented polymer products.

Wyeth patented his process in 1973 and though recycling was not an avid idea that that time, the first PET soda bottle went into recycling soon in 1977.[13] Today, recycled PET is widely used as synthetic fiber with a major part of it used in making polyester carpets: nearly half of the polyester carpet made in the U.S. today come from recycled PET bottles.[15]

RICHARD E. MORLEY



Richard Morley was an American mechanical engineer and is considered one of the 'fathers' of the programmable logic controller (PLC). Morley designed the first PLC with his team Mike Greenberh, Jonas Landau and Tom Boissevain and called it 084, as it was their 84th project at Bedford Associates.[16] The introduction of PLCs kicked-off the 3rd industrial revolution, leading to the development of an entire industry of digital control solutions.

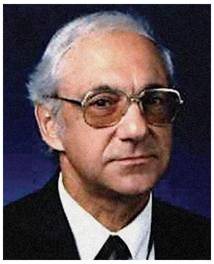
In 1964, when Morley was unemployed and working in uninteresting design jobs, he decided to pursue his interest in engineering by starting his own consulting firm with his friend Geogre Schwenk under the name Bedford Associates. Initially, they worked with machine tool firms to help them transition into solid-state manufacturing. Eventually, Morley realized that the projects he worked on were similar and work became monotonous. He decided to use his creativity and engineering acumen to invent a controller that would automate industrial processes with multiple input/output arrangements in real-time and replace hard-wired relay controls.

During those times, manufacturing facilities were operated by relay control systems. Control rooms were large with walls full of relays, terminal blocks and wired connections. The main

challenges were a lack of flexibility to make process changes and the extensive time required to adjust these changes. Morley managed to design the functions of a PLC that offered advantages of uninterrupted processing, flexibility, fast reaction time and direct mapping into memory—revolutionizing manufacturing process control. The PLC was designed to be robust under severe temperature and moisture conditions and used large metal fins to transfer out air, keeping electronics dirt free. The product was capable of operating as a modular digital controller and was hence named Modicon, a brand now owned by Schneider Electric.[17]

Morley has been widely recognized in numerous publications and awards from the International Society of Automation, Instrumentation, Systems, and Automation Society, the Franklin Institute, the Society of Manufacturing Engineers and the Engineering Society of Detroit.[18] He was also inducted into the Manufacturing Hall of Fame. The Society of Manufacturing Engineers offers the Richard E. Morley Outstanding Young Manufacturing Engineer Award for outstanding technical accomplishments in the manufacturing profession by engineers aged 35 and under.[18]

ODO JOSEF STRUGER

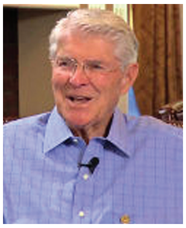


Odo J. Stuger, an Austrian engineer and scientist, is recognized as the pioneer of modern-day automation and shares the credit as 'father of PLC' alongside Richard Morley. During 1958–1960,[19] Struger led his engineering team at Allen-Bradley in developing the programmable logic controller and also coined the acronym PLC for programmable logic controllers. Allen-Bradley became the pioneering leader in programmable logic controllers in the U.S., and PLC remains a registered trademark of the Allen-Bradley Company (now Rockwell Automation).

Struger's work on PLCs was built upon the concepts studied in his doctoral research in "The process for quantitative handling of positioning errors in numerical control machines," at Vienna University of Technology.[19] His invention proved to be a 'rugged industrial computer' that, through precise numerical control of machinery, soon became ubiquitous in manufacturing environments across the world.

Struger was born in Carinthia, Austria, and studied at the Vienna University of Technology. In 1958, he moved to Milwaukee, Wisconsin (U.S.) to work as a research engineer at Allen-Bradley. Struger grew within the company and held the position of Vice President of technology until retirement in 1998.[19] He was associated with the development of the National Electrical Manufacturers Association (NEMA) standard for PLCs and IEC 1131-3 programming language standard. Struger has 50 patents to his credit in the U.S. and Canada. He received the Prometheus Award in 1996, authored more than 40 technical papers and is an inductee to the Automation Hall of Fame at the Chicago Museum of Science and Industry. To honor Struger's legacy, Rockwell Automation established the Odo J. Struger Automation Award for future engineers' exceptional advancements in the control and automation fields.[20]

JOHN MOONEY AND CARL D. KEITH



John Mooney, an American chemical engineer, and Carl D. Keith, a chemist, created the three-way automotive catalytic converter while working at Engelhard Corporation in 1973 and solved a major environmental problem—making automobile exhaust 98% cleaner.[21] An EPA report recognized this invention as one that helped save 100,000 lives and prevent many more cases of lung and throat ailments.[22] Today, catalytic converters are the key-emissions control components in automobiles worldwide.

The earliest catalytic muffler was developed by Eugene Houdry as a generic device that could convert carbon monoxide (CO) and unburned hydrocarbons (UHCs) from automobile and industrial exhausts. Houdry launched his company Oxy-catalyst and his catalytic converter design was patented in 1962.[23] However, fuel still contained tetraethyllead (TEL) as an anti-knock agent, which poisoned the catalyst in the converter. It took the passing of the Clean Air Act in 1970 and the ban of TEL for converters to be recognized and become a piece of standard equipment in automobiles.

Mooney and Keith, while working at Engelhard Corporation (acquired by BASF in 2006)[24], developed the three-way catalytic converter, where the exhaust gas components are UHCs, and CO are oxidized and nitrogen oxides (NO_x) are reduced to water, nitrogen and carbon dioxide (CO_2). The inherent complexity of the reaction implied the need for a bulky two-stage converting system. Mooney, however, theorized that if the fuel-to-air ratio was correct, the exhaust would provide just the right amount of oxygen for a one-stage converter to treat all three pollutants at once.

Equipped with his idea and a 'can-do' attitude, Mooney garnered his supervisor Keith's support to allow him to convince auto manufacturers to include an oxygen sensor to their engines. The sensor was intended to monitor fuel-to-air ratio at a level where the one-stage converter could function successfully. Volvo agreed to the proposal and soon the sensors were successfully incorporated into other automobiles, as well. The converter is a small can-shaped device that installs at the exhaust pipe under vehicles. A combination of rare-earth oxides and base metal oxides along with platinum and rhodium were used together in the catalyst. The engine exhaust passes over a specialized honeycomb-shaped structure, where a washcoat of catalyst materials acts as active sites for reactions. The design ensured an adequate amount of oxygen was offered for the oxidation and allowed all three pollutants to be targeted at once.

Both Keith and Mooney received the 2001 Walter Ahlstrom Prize and earned the National Medal of Technology in 2002 for their invention.[24] Engelhard (now BASF) continues to lead the development of automotive emissions catalysts.

ACKNOWLEDGEMENTS

Hydrocarbon Processing would like to thank several institutions/companies/websites for the use of archived images of industry pioneers. These include the National Inventors Hall of Fame, MIT Lemelson, Science History and The National Academy of Engineers.

LITERATURE CITED

Complete literature cited available online at www.HydrocarbonProcessing.com.

J. BUEHLER, Petrogenium, Buehler Consulting, Houston, Texas

Overview of decarbonization pathways for the oil and gas and petrochemical industries—Part 2

This article is the continuation of an article detailing the seven pathways to decarbonizing the oil and gas and petrochemical industries. Part 1, published in the May issue of *Hydrocarbon Processing,* covered sustainability and reviewed the following three pathways: green and blue hydrogen (H_2); biofuels, renewable fuels and e-fuels; and the circular carbon cycle. This article will discuss the first four pathways to decarbonizing the oil and gas and petrochemical industries:

1. Energy efficiency/ stopping methane leakage
2. New technologies
3. Electrification
4. Carbon capture, utilization and storage (CCUS) technology.

Additionally, as detailed in Part 1, renewable power is also key to generating green H_2 by electrolysis.

Pathway 1: Energy efficiency/stopping methane leakage.

Improving energy efficiency and stopping methane leaks is the cheapest way to reduce CO_2e emissions and CO_2e intensity. If plants do not burn or leak the fuel to atmosphere, CO_2e is not produced and the organization saves on the cost of the fuel (i.e., a win-win situation).

Some energy projects will have a negative cost of CO_2 capture, after considering the capital expenditure cost, and the variable and operational costs. Capital energy projects can have less than a 2 yr–3 yr payback, while maintenance projects can achieve payback in just 2 mos–3 mos.

Conduct an energy and greenhouse gas (GHG) study at your plant. Operators should use their internal company specialists, or an outside energy process consultant, to obtain an independent review of the facility for a fresh look toward identifying any possible overlooked energy/GHG reduction opportunities. This study should be structured using process unit-specific checklists, and should identify opportunities, conduct a high-level estimate of potential energy/GHG savings, estimate the cost of projects, and develop a prioritized list of energy/GHG improvement opportunities. Quick-win projects, along with projects that could be accomplished during the next turnaround, should be flagged for early implementation. Savings from quick-payback projects can be used to fund additional energy projects.

An energy and GHG reduction study should focus on the following items: process optimization, equipment performance and utility system optimization.

Process optimization. Process optimization includes the following focal points:

- **Distillation:** Optimize pumparounds, avoid over refluxing, utilize advanced process control to meet specifications and minimize product giveaway, and optimize column pressure and feed temperature.
- **Reactions:** Optimize H_2:oil ratios and yield vs. energy vs. run length, and catalyst selection.
- **Heat recovery:** Use pinch analysis to identify opportunities to recover process heat lost to air or cooling water. Use a heat pump to upgrade low-temperature waste heat to a useable heat source—this process can recover 2–3 times the energy required by the heat pump. Use an organic Rankine cycle to convert waste heat to electricity.
- Stop methane leakage by conducting surveys and fixing leaks. The use of drones can help with inspection work.
- Use online, real-time monitoring of energy key performance indicators (KPIs) available to operators. Monetize the financial gap between the current operation and the previously demonstrated best performance. Operators should use advanced process control, optimizers and digital twins.
- Stop continuous flaring, and revise procedures to minimize flaring during startups and shutdowns.

Equipment performance. The following steps can be taken to enhance equipment performance:

- Improve fired heater efficiency by monitoring stack temperatures and excess air, adjusting burners, cleaning the convection section and repairing air preheaters. A 2% oxygen reduction in stack flue gas equals a 1% efficiency improvement. A 20°C reduction in stack temperature (with reasonable excess air) equates to a 1% efficiency improvement.
- Monitor vacuum improvements on condensing turbines. An improvement of 1.2 in. Hg (0.6 psia) equates to a 5% improvement. On a large 250,000-lb/hr condensing turbine, a 5% improvement saves more than $850,000/yr in fuel cost (natural gas at $6/MMBtu, boiler efficiency of 85%).

- Calculate steam turbine isentropic efficiency (steam temperature in vs. temperature out). Consider replacing small, inefficient (30% efficiency) steam turbines with motor drives.
- Minimize compressor recycling by using suction throttling, inlet guide vanes, variable frequency drive motors, reciprocating compressor variable pocket unloaders and re-rotor compressors. In addition, personnel should check the inlet temperature to the compressor stages and clean interstage heat exchangers.
- Optimize the operation of parallel compressors to avoid having excess recycle—this can occur in LNG and gas processing plants.

Utility system optimization. The following recommendations can help optimize a plant's utility system:

- Utilize a steam header ladder diagram to obtain an overview of steam demand at each pressure level, which utility boilers and turbines are operating, and steam letdowns (high pressure to medium pressure to low pressure). In addition, optimize the use of steam letdown turbines to recover mechanical energy lost through pressure control letdown valves.

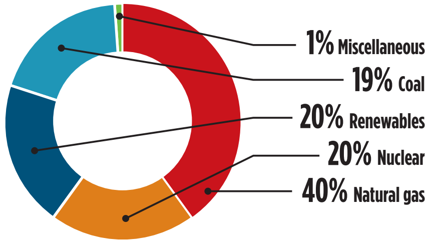


FIG. 4. U.S. power generation by source, 2020. Power generation totaled 4.12 T KWh. Source: U.S. EIA.

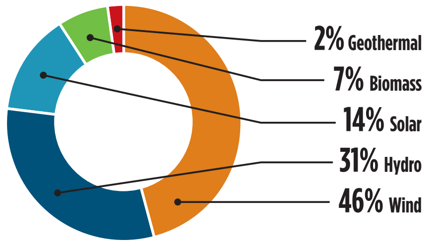


FIG. 5. U.S. utility-scale electricity generation by renewable sources, 2021. Total renewable power generated was 826 B KWh. Source: U.S. EIA.

- Eliminate low-pressure steam venting by replacing pump turbine drivers exhausting to the low-pressure steam header with motor drives. If a motor is in critical service, keep a spare turbine on hot standby with its casing and inlet/outlet lines hot.
- Remember that a steam trap/leak program can pay for itself in less than 6 mos. Operators should also fix air header/air compressor leaks and shut down rental diesel air compressors.
- Consider installing a cogeneration unit if more steam is needed or if old boilers need to be replaced.

Pathway 2: New technologies. When it is time for a major maintenance turnaround or debottlenecking project, plant owners should consider using the latest technologies. This can range from re-rotoring compressors and upgrading turbines to using higher-capacity trays in columns or higher-efficiency heat transfer equipment. The following is a partial list of technology items to consider:

- For new grassroots projects, seek new technologies that provide higher yields and reduced energy consumption, instead of replicating an existing, older, less-efficient design.
- For revamps or new units, use new improved catalysts to increase yield and reduce recycle, thus reducing energy and GHG intensity.
- For the electrification of compressor drivers on offshore platforms, use offshore wind or onshore green power.
- Install once-through steam generators (OTSGs) on open-cycle gas turbines to recover gas turbine exhaust heat. The light weight and small footprint of OTSGs allow for retrofitting.
- Use a divided wall column (which uses one distillation column instead of two columns) for reduced capital and operational expenditures.
- Install heat pumps on columns separating close boiling temperature components (e.g., ethylene/ethane splitter, propylene/propane splitter).
- Using a low-emissions cracking furnace designed to shift more fired duty into process heating in the firebox and generate less steam in the convection section, can reduce fuel requirements by 30%. Reduced steam production provides the opportunity to replace condensing steam turbines with efficient motors using renewable power.
- Use new monoethylene glycol (MEG) technology, such as a proprietary MEG process[b] that reduces steam usage by 20%.
- Utilize new propylene oxide (PO) processes, such as a proprietary PO process[c] that reduces energy usage by 35% and wastewater by 70% vs. existing PO technologies.
- Use new ethylene processes (e.g., methanol-to-olefins, oxidative coupling methane).
- Utilize the Allam-Fetvedt cycle for power generation (e.g., an oxy-fired gas turbine using a supercritical CO_2 cycle, where CO_2 offtake is ready for CO_2 pipelines). This creates clean power with no net carbon to atmosphere.

Pathway 3: Electrification with renewable power. In 2020, electrical generation in the U.S. accounted for 32% of CO_2 emissions from the burning of fossil fuel. Electrification of process equipment and transportation vehicles using renewable power will be critical to reducing CO_2 emissions. Renewable power has no CO_2 footprint, and includes solar photovoltaic (PV) systems, as well as concentrated solar, onshore and offshore wind, hydroelectric and geothermal systems. Nuclear power also has no CO_2 footprint. One MWh of electricity will power approximately 200 homes on a hot summer day.

According to the Center for Climate and Energy Solutions, approximately 29% of global electricity generation was produced via renewables in 2020. Also in that year, electric power in the U.S. was generated from natural gas (40%), renewables (20%), nuclear (20%), coal (19%) and miscellaneous (1%)—and that year was also the first year in which renewable power constituted a larger market share than coal power (**FIG. 4**). In 2021, most of the installed utility-scale power generation in the U.S. from renewable sources was wind (46%), followed by hydro (31%), solar

(14%) and biomass (7%) (**FIG. 5**). Future installed power generation will be predominately solar and onshore/offshore wind. Renewable power can be used in an electrolyzer to convert water to green H_2 and oxygen. H_2 is another key decarbonization pathway, which was discussed in Part 1 of this article.[21]

Renewable power is not available at a constant supply rate, so power grids must be upgraded to accept fluctuations in renewable power supply. Solar is not available at night, and is reduced on cloudy days. Wind also varies between day and night, and from summer to winter. Hydro varies with the seasons. Over a year, the utilization of installed capacity for solar was 20%–28%, for wind 35%–40% and for hydro 35%–45%.[22]

Energy storage is required to stabilize the grid and maximize the use of renewable energy. Batteries can store excess renewable power during the day and then release this power into the grid at night. Significant research in battery energy storage systems is ongoing. Challenges include the amount of power that can be stored (which is limited by battery size and weight), and the amount of the dispatch rate of the power back into the grid when there is sudden demand. Lithium-ion batteries are most used in electric vehicles (EVs). Flow batteries are an emerging technology being developed for storing renewable energy. Flow batteries contain a water-based electrolyte liquid that flows between two separate tanks within the battery. When charged, a chemical reaction occurs that enables the energy to be stored and subsequently discharged.

In addition to batteries, other methods for storing excess renewable power when grid demand is reduced include the following:

- Compressing air in underground storage domes for later use to drive turbines
- Pumping water to elevated storage, to be used later to drive hydropower turbines
- Utilizing molten salts for thermal storage
- Using electrolyzers for green H_2 generation.

It will be imperative to upgrade and expand the U.S. electrical grid to supply remotely generated renewable power to urban areas that need power. To minimize line loss (which can be 2%–4%), high-voltage direct current cables can be used. Reduced transmission line loss will decrease capital expenditures and improve efficiency. Smart power grids will provide real-time knowledge of power generation status, along with consumption balance and the impact of this power on the grid. They will also enable operators to instantly optimize available resources to stabilize the grid.

The cost to install solar and wind generation technologies has dramatically reduced. Capital expenditures for utility-scale solar and wind power are competitive with gas-fired combined-cycle power plants. Utility-scale renewable power has a large footprint requiring a lot of land. Solar PV systems need approximately five acres of land to generate 1 MW of electricity (18% efficiency).[23]

A land-based wind farm would require approximately 60 acres to generate 1 MW

TABLE 1. CO_2 reduction by using renewable power vs. generation from fossil fuels

	Kg CO_2/ MWh generated	CO_2 for 100-MW power plant, tpy
Modern combined-cycle power plant (64% efficiency)	283	248,000
Natural-gas-fired, open-cycle peaking plant at 34% efficiency	532	466,000
Coal-fired power plant (100 kg CO_2/1 MMBtu coal) at 34% plant efficiency	1,040	911,000

of power. Only 3% of the land is occupied by turbines, leaving the rest of the land for farming and grazing. The layout of the wind turbines is critical to avoid secondary air from nearby turbines.[24] Land-based wind turbines presently have a capacity of 3 MW, which will increase to 6 MW as technology improves. With more favorable winds, offshore wind turbines will have a higher capacity that can be scaled up to 17 MW by 2030.[25,26] An offshore wind turbine can have a rotor diameter of 170 m. Large offshore wind farms are being developed in the North Sea and are planned for installation off the northeast coast of the U.S. (i.e., Massachusetts, New Jersey and New York). One beneficial factor is that offshore oil/gas platform technology can be applied to offshore wind. Integrated oil and gas companies are investing in both solar and wind generation.

Solar PV systems and wind turbines are unable to supply all the green power required, so there will still be a role for nuclear power after 2035. Development work on the next generation of nuclear reactors is ongoing. For example, NuScale is developing a small modular nuclear reactor, which is a light water reactor that does not need pumps for circulating reactor cooling water. TerraPower is developing its Natrium reactor, which is a sodium-fast reactor combined with molten sodium salt heat storage.

Oil companies are moving into the generation and direct sale of renewable electricity to customers. Electrification of transportation vehicles—such as cars, trucks and trains—is a key pathway for reducing Scope 3 emissions resulting from the burning of gasoline and diesel. In the U.S., the Biden administration has set a goal for half of all new vehicles sold in 2030 to be zero-emissions vehicles, including battery electric, plug-in hybrid electric, or fuel cell electric vehicles (FCEVs). By 2035, General Motors plans to stop making internal combustion cars. The installation of electric charging stations along major highways, and at homes and workplaces, will be necessary to achieve these goals—and accomplishing this will constitute a major capital expenditure.

According to the U.S. Environmental Protection Agency (EPA), 1 gal of gasoline contains 33.7 kw of energy and, when burned, generates 8.87 Kg CO_2.[27] If one person drives 20,000 mi/yr with a 35-mpg car, they will generate 5 tpy of CO_2. If a subdivision has 200 EVs using renewable power, CO_2 can be reduced by 1,000 tpy. **Note:** If green power is generated and converted to H_2 for use in an H_2 FCEV, CO_2 is reduced by only 42% vs. using green power directly in a battery EV. A battery EV will average 3.5 mi/kWh. If a person drives 35 mi, 10 kWh of electric power is used. If power costs \$0.13/kWh, this equates to \$1.30. Driving 35 mi in a gasoline vehicle would use 1 gal of gasoline at a cost of approximately \$3/gal. In this example, using an EV would save the driver \$1.70.

TABLE 1 shows the CO_2 reduction impact of using renewable power vs. three current methods of electrical generation. Using 100 MW of renewable power will reduce CO_2 emissions by 248,000 tpy vs. a combined-cycle power plant; 466,000 tpy vs. a natural-gas-fired, open-cycle gas turbine peaking plant; and 911,000 tpy vs. a coal-fired plant. The impact of phasing out coal and open-cycle natural gas plants (which have no gas turbine heat recovery) with renewable power is clear. Blending 20% H_2 (vol%) into natural gas as fuel for a combined-cycle plant will reduce CO_2 emissions by approximately 7%.

Industrial opportunities for electrification to reduce Scope 1 and Scope 2 CO_2e emissions include:

- Switching condensing turbines to electric drives, considering the impact on steam and fuel balances
- Replacing old and inefficient steam turbines that cause low-pressure steam venting with electric drives
- Using electrically driven heat pumps to upgrade unusable low-level heat to a higher usable level
- Switching medium-temperature fired heaters to electric power, and switching out larger heaters to use green H_2 as a fuel
- Replacing open-cycle gas turbine drives in LNG plants with large electric drives
- Replacing the gas turbines on offshore platforms with electric motors that use renewable power from offshore wind turbines and land-based renewable power
- Electrifying ethylene plants, using motor-driven cracked gas and refrigeration compressors.

Ethylene crackers are a large source of CO_2 emissions. Replacing the large thermally inefficient condensing steam turbines with electric drives reduces the need for steam generation. The cracker furnaces are the primary source of CO_2 emissions in an ethylene plant. Adding air preheat and redesigning the furnace to recover more of the heat in cracking coils and to produce less steam in the convection section can reduce furnace fuel usage by 30%.[28] Work is ongoing to replace steam cracker furnaces with electrically heated furnaces. Earlier this year, pilot-scale testing began of a roto dynamic reactor that could replace the cracking furnace. The roto dynamic reactor uses electrically powered rotor blades that transfer mechanical energy into thermal energy in just nanoseconds, thus reducing the reactor residence time—this can, in turn, increase ethylene yield.[29]

Renewable electricity can also be used to produce green H_2, which can be used to hydrogenate biofeedstocks to produce biofuels. Green H_2 can also be used to make synthetic fuels or e-fuels with a low carbon footprint through the e-chemistry of Power-2-X (P2X) conversion technology. P2X technology reacts green H_2 with CO_2 to produce gaseous fuels [H_2, methane (CH_4)] or liquid fuels [methanol, synthetic fuels, sustainable aviation fuel (SAF)] and chemicals [ammonia (NH_3)]. The reaction of green H_2 with captured CO_2 to produce syngas for use in Fischer-Tropsch chemistry to make e-fuels was discussed in Part 1 of this article, which was published in the May issue of *Hydrocarbon Processing.*

Pathway 4: CCUS technology. Another pathway to decarbonization is through CCUS. This technology plays a key role in decarbonizing H_2 produced by steam methane reforming (SMR) to produce blue H_2. Carbon capture is also key to capturing CO_2 in processes (such as natural gas, ethylene oxide, ethanol and H_2 production) that produce streams of high CO_2 concentration. Additionally, CCUS is instrumental in reducing CO_2 emissions in the hard-to-decarbonize steel and concrete industries.

The oil and gas industry is a leader in developing and deploying CCUS technology. In 2020, global anthropogenic (human-made) CO_2 emissions totaled 34 gigatons, with only 40 megatons (0.12%) being captured.[30] Oil and gas operations captured 70% of the CO_2 emitted. The oil and gas industry has the capability to implement CCUS on a large scale. Depending on the success of implementing the other decarbonization pathways, the International Energy Agency (IEA) estimates that CCUS may be needed to capture as much as 7.6 gigatons/yr of CO_2 to meet 2050 global net-zero goals.

In the U.S., the 45Q tax credit provides an incentive to install CCUS at \$35/t for CO_2 used for enhanced oil recovery (EOR) and at \$50/t for CO_2 that goes into storage. These incentives are not large enough for capturing CO_2 in diluted CO_2 streams. Depending on oil prices, drillers may pay \$20/t–\$30/t for CO_2 used in EOR. CO_2 capture cost ranges from \$30/t–\$40/t for streams with high CO_2 concentrations and can cost \$60/t–\$90/t for CO_2 captured from fired heater flue gas, where CO_2 content is only 8%–10%.

The cost to capture CO_2 in natural-gas-fired power generation ranges from \$60/t–\$90/t. Adding carbon capture and storage (CCS) to a combined-cycle power plant derates net power output by 12%–15%. Derating is the result of the heat required to regenerate CO_2 absorber solvent (2 Gj/t CO_2) and power for CO_2 compression (200 kWh electricity/t CO_2). Note that 2 Gj/t CO_2 is roughly equivalent to 1 t steam/1 t CO_2.[31,d] There are three primary methods of CO_2 capture from combustion:

- **Post-combustion:** This method includes burning fuel (CH_4 is used in examples) and capturing CO_2 in a fired heater stack. Flue gas contains 8%–9% CO_2, 70% nitrogen, 18% water and 2%–3% oxygen at atmospheric pressure. **Note:** A gas turbine's exhaust without duct burner firing contains only 3%–5% CO_2. Post-combustion capture uses chemical solvents, such as hindered amines. Net CO_2 removed from flue gas capture is about 90% of the actual CO_2 captured, since absorber solvent regeneration and CO_2 compression have a CO_2 footprint that reduces the actual CO_2 removed from the atmosphere.
- **Oxy-firing system:** This system involves burning CH_4 fuel with oxygen (supplied from an air separation unit) in a fired heater. Flue gas contains CO_2 and water but no nitrogen. No CO_2 capture unit is required. Compressing flue gas removes water to produce a pure CO_2 stream. Personnel must seal leaks in the convection section to keep it from sucking in air, which will add nitrogen in the flue gas, leading to the captured CO_2 being off specifications.

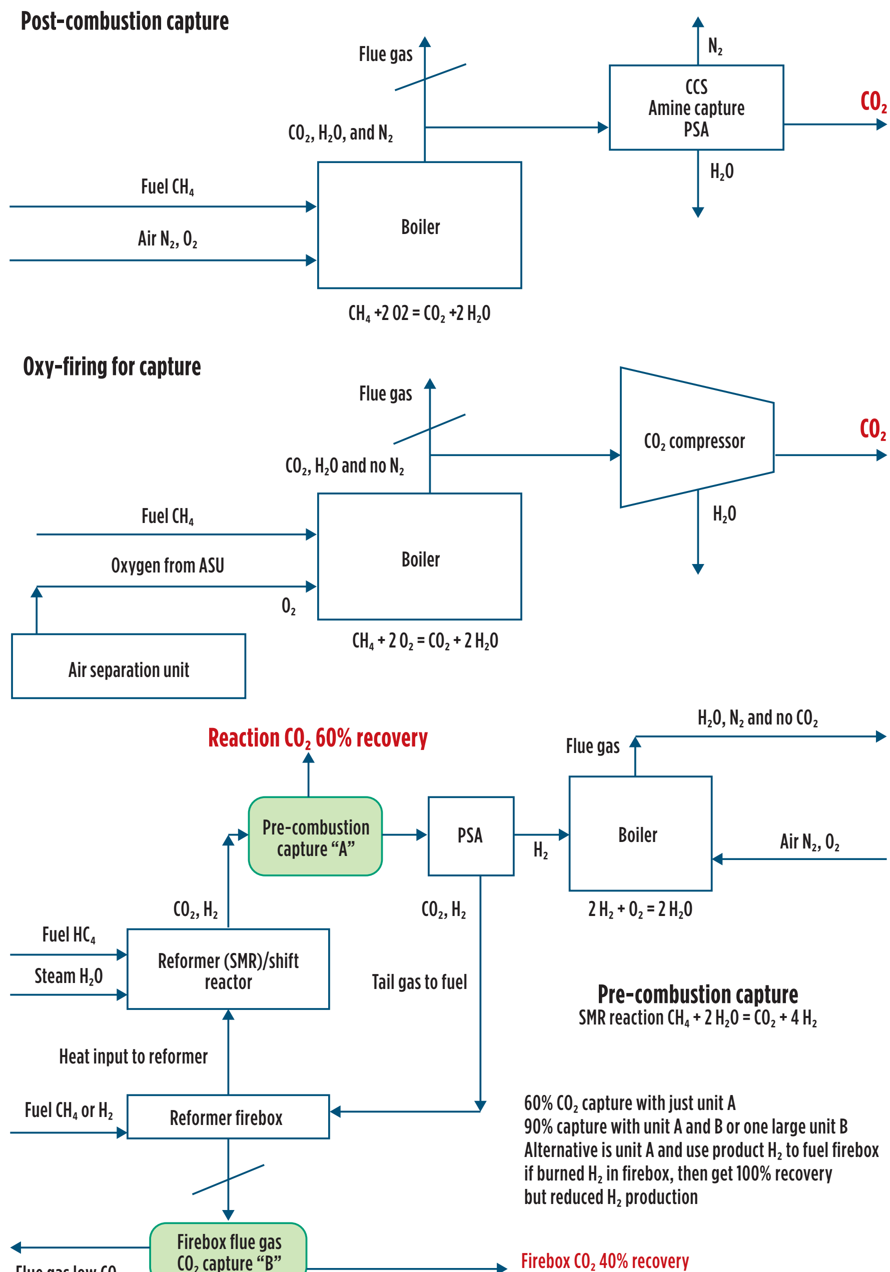


FIG. 6. Three methods to capture CO_2 from combustion.

- **Pre-combustion:** If H_2 is burned in a fired heater, the flue gas contains no CO_2; therefore, no CO_2 capture is required for the heater. The fuel is sent to a reformer, followed by a water-gas shift reaction to convert the fuel into H_2 and CO_2. The CO_2 is then captured at the reformer outlet (higher % CO_2 and higher pressure makes CO_2 easier to capture than from flue gas) by using a physical solvent in an absorber (see "A" in **FIG. 6**). The process, with the CO_2 removed, proceeds from the reformer's outlet to a pressure swing adsorption (PSA) unit that recovers pure H_2. The PSA tail gas goes to the fuel system. Reforming is an endothermic reaction requiring heat. In an SMR unit, 60% of CO_2 is generated by the reformer/water-shift reaction, and 40% is from the reformer's firebox, which generates the heat required for the reaction. The CO_2 absorber on the reformer's outlet will recover approximately 60% of the produced CO_2. To recover the remaining amount, a second CO_2 absorber (marked "B" in **FIG. 6**) on the firebox flue gas stream would need to be added to obtain a 90% CO_2 capture. An alternative is to fuel the firebox with H_2—which produces no CO_2 but reduces H_2 production. Installing a single, large CO_2 absorber on the firebox flue gas (B) may be more cost effective. An alternative to the reformer is to use an auto thermal reformer (ATR) to produce H_2. An ATR oxidizes (burns) part of the CH_4 feed in the reactor to generate the required reaction heat, thereby eliminating the need for a firebox. The ATR uses oxygen from an ASU as an oxidant, so there is no nitrogen in the outlet of the reactor. The remainder of the ATR outlet is H_2 and CO_2, which can be separated like a basic reformer, using a CO_2 absorber or a PSA unit. Using partial oxidation of methane with carbon capture is another route to produce H_2 for fueling a fired heater. **FIG. 6** shows a high-level schematic of the three methods to capture CO_2 from combustion.

CO_2 is captured and removed from process streams by either chemical absorption or physical separation. Chemical absorption involves a chemical reaction between the amine solvent and CO_2, and requires the regeneration of the solvent—1 t steam/1 t CO_2, which is equivalent to 2 Gj/t CO_2. Physical separation can include either adsorption, absorption, cryogenic separation, or dehydration and compression. Physical adsorption uses solid surfaces like zeolites, metallic oxides, alumina or activated carbon. After the CO_2 is captured by the physical adsorbent, the CO_2 is then released by cycling temperature in temperature swing adsorption (TSA), pressure in a PSA unit, or vacuum in vacuum swing adsorption. Physical absorption utilizes liquid solvents that are regenerated by flashing the solvent to a lower pressure in the regenerator, which does not require steam, thus resulting in less energy being used.[32]

Significant research is being conducted to reduce CO_2 capture costs in CCUS. The U.S. Department of Energy's Carbon Negative Shot initiative seeks to reduce the cost of CO_2 removed from the atmosphere to less than \$100/t by 2030 via either direct air capture or by helping forests, agriculture and energy crops capture and store CO_2. Additional ongoing CCUS research includes:

- Absorption with chemical solvents (hindered amines) to lower costs by reducing the energy required in the solvent regenerator for aqueous solvent regeneration
- Absorption with physical solvents, which can be regenerated with vapor flashing and without requiring steam (resulting in lower energy usage)
- Utilization of other non-aqueous or water lean solvents to reduce the energy required for regeneration
- Adsorption onto solids, hydrated solid sorbents and metals with thermal or vacuum swing adsorption (solid adsorbents can withstand higher gas temperatures than solvents and can avoid solvent degradation)
- Direct air capture, as the low CO_2 concentration in air at 420 ppm (0.042%) vs. flue gas at 9% is a challenge (direct air capture requires processing 214 times larger volume of air to remove the same amount of CO_2 as flue gas capture, requiring large air contactors, like cooling tower units with air flowing up through packing where it contacts a solvent or solid adsorbent)
- Cryogenic separation of CO_2 from H_2 on the outlet of a steam methane reformer
- A process for growing algae to absorb CO_2 from air (photosynthesis), and then converting algae oil to biofuels.

Large-scale CCUS installations using chemical solvent absorber technology are proven to be effective with 1-MMtpy to 3-MMtpy CO_2 capture trains [e.g., Boundary Dam in Saskatchewan, Canada; Petra Nova in Houston, Texas (U.S.); QUEST in Alberta, Canada on a H_2 steam methane reformer (1.1 MMtpy); and LaBarge, Wyoming (U.S.) in a sour gas treating plant (6 MMtpy). CO_2 must be compressed to 100 bar for pipeline and underground storage. To prevent carbonic acid corrosion in CO_2 pipelines, there are tight specifications on oxygen at 50 ppm and on water at 20 ppm. Because of the high capital expenditures of CO_2 capture units, pipelines and CO_2 storage areas, a "hubs and clusters" concept is necessary to obtain sufficient CO_2 volumes for an economically efficient project.

There will be a need for regional CO_2 pipelines to collect CO_2 and then to either provide it as a feedstock or deliver it to storage. This activity will see competitor companies working together to jointly reduce their CO_2 emissions. Several CCS pipeline projects have been announced in the UK and near Rotterdam/North Sea. In the U.S., discussions are ongoing for a CCS network to be developed in the Houston Ship Channel to capture up to 100 MMtpy of CO_2 by 2040 and to store it offshore in the Gulf of Mexico. Additionally, Talos Energy has announced plans to develop CO_2 pipelines and storage along the Mississippi River corridor in Louisiana and on the Texas Gulf Coast.

CO_2 storage can include underground saline reservoirs, depleted oil and gas reservoirs, porous rock, subsea or basalt. When CO_2 is stored, the operator must measure, monitor and verify the

CO_2. No CO_2 leakage can occur to atmosphere or into adjacent underground structures. For any new grassroots facility, a study should be done to identify potential nearby CO_2 storage locations.

Some uses for CO_2 include EOR (which uses 0.3 t CO_2/bbl–0.6 t CO_2/bbl), along with processes for freeze drying of food, carbonization of beverages, production of urea to make fertilizer and enhanced photosynthesis in greenhouses. Growing biofuel feedstocks (energy crops) are an effective way to remove CO_2 from the atmosphere and to produce low-carbon biofuels.

Captured CO_2 can be reacted with green H_2 made from renewable power to produce synthetic fuels with a low CO_2 footprint (i.e., P2X or CO_2 e-chemistry).

Nature captures CO_2 from the atmosphere by photosynthesis, which converts CO_2 and water to oxygen and glycose by electron transfer. Research is ongoing to develop electrochemical cells that will replicate and accelerate this electron transfer by using electrolysis to produce synthetic fuels and chemicals. In an electrochemical cell, a catalyst embedded in the cathode will reduce CO_2 to carbon monoxide or other products on one side of the cell. On the other side, the anode oxidizes water from the electrolyte to produce oxygen. This is a promising path for using renewable power and captured CO_2 to produce chemicals and fuels.[32]

Takeaways. The oil and gas industry is rebranding itself as an energy provider, and is transitioning to selling low-carbon-intensity energy products that include renewable wind and solar power, green and blue H_2, low-carbon LNG, biofuels, renewable diesel, e-fuels, e-gasoline and SAF.

Improving energy efficiency and stopping methane leakage are the cheapest ways to reduce CO_2 emissions. For new facilities or major plant expansions, project developers should consider incorporating new technologies that have better energy efficiency and lower carbon intensity. For major turnarounds requiring catalyst changeouts, operators should consider newer formulated catalysts for higher yields and lower energy intensity.

The pathways to reduce Scope 1 and Scope 2 CO_2 emissions include:

1. **Energy efficiency/stopping methane leakage:** Improving efficiency in existing and new facilities by maintaining energy recovery equipment, stopping or minimizing routine flaring (such as minimizing flaring on startups and shutdowns), and identifying and stopping methane leakage
2. **New technologies:** Using new process technologies and catalysts to improve yields and reduce energy intensity
3. **Electrification:** Electrifying process equipment; using renewable electric power (e.g., wind, solar, hydro and nuclear power); and investing in battery storage, and in using and selling renewable power
4. **CCUS:** Incorporating cost-effective CCUS technology, joining regional CCUS networks and using captured CO_2 in e-chemistry to produce e-fuels
5. **Green and blue H_2:** Producing, utilizing and selling low-carbon H_2, as well as using H_2 as an energy carrier and to produce e-fuels.

Pathways to reduce Scope 3 emissions for the oil and gas industry include producing and selling energy products with low-carbon intensities, such as:

6. **Biofuels, renewable fuels and e-fuels:** This includes producing renewable diesel, SAF, synthetic fuels, e-gasoline, low-carbon LNG and renewable electricity, as well as green and blue H_2.

The seventh pathway will help reduce Scope 3 emissions from petrochemicals production:

7. **Circular carbon pathway:** This pathway includes recycling plastics by either mechanical reprocessing or chemical means (pyrolysis, gasification); using renewable feedstocks like bio-naphtha, hydrogenated vegetable gasoil (diesel) and plastic pyrolysis oil to produce the base chemicals ethylene and propylene; and producing synthetic chemicals, such as ethanol and methanol, by using renewable H_2 and captured CO_2.

The good news is that the oil and gas and petrochemical industries have the technology and assets needed for offshore wind turbines, blue/green H_2 production, and CO_2 capture and storage. They also have the refinery units and technology to produce renewable fuels. These industries are prepared for the journey to complete this crucial energy transition to a lower-carbon world. HP

NOTES

[b] Shell's OMEGA process

[c] BASF and Dow Chemical Co.'s hydrogen peroxide to PO (HPPO) technology

[d] According to Buehler Consulting, a 15% reduction is based on 2 Gj/t CO_2 to regenerate solvent, and on 200 kWh electricity/t CO_2 for compression

LITERATURE CITED

21. U.S. Energy Information Administration (EIA), "What is U.S. electricity generation by energy source?" online: https://www.eia.gov/tools/faqs/faq.php?id=427&t=3
22. Statista, "Capacity factors for selected energy sources in the United States in 2020," online: https://www.statista.com/statistics/183680/us-average-capacity-factors-by-selected-energy-source-since-1998/
23. Yacoubou, J., "Solar farm land requirements: Top 7 tips for farmers, ranchers, and landowners," GreenCoast, December 5, 2021, online: https://greencoast.org/solar-farm-land-requirements/
24. Gaughan, R., "How much land is needed for wind turbines?" Sciencing, May 10, 2018, online: https://sciencing.com/much-land-needed-wind-turbines-12304634.html
25. Patel, S., "Changing winds: Emerging wind turbine technologies," *Power,* June 1, 2021.
26. Proctor, D., "Winds of change revitalize West Texas," *Power,* September 1, 2021.
27. U.S. EPA, "Code of Federal Regulations: Part 98," June 13, 2017.
28. Boekel, T. V. and P. Oud, "Low-emission ethylene furnace," AIChE Spring Meeting and Global Congress on Process Safety, April 19, 2021.
29. Coolbrook, "Coolbrook makes petrochemical industry more sustainable with innovative pilot," July 28, 2021, online: https://coolbrook.com/news/2021/coolbrook-makes-dutch-petrochemical-industry-more-sustainable-with-innovative-pilot/
30. IEA, "Net zero by 2050: A roadmap for the global energy sector," May 2021, online: https://iea.blob.core.windows.net/assets/deebef5d-0c34-4539-9d0c-10b13d840027/NetZeroby2050-ARoadmapfortheGlobalEnergySector_CORR.pdf
31. IEA, "Energy Technology Perspectives 2020: Special report on carbon capture utilization and storage," October 19, 2020.
32. Boerner, L., "How can we convert CO_2 from threat to asset?" *Chemical and Engineering News,* October 11, 2020.

JACK BUEHLER is an Energy/GHG Consultant. He recently retired from Shell as a Principal Engineer identifying energy and GHG reduction opportunities for existing assets and new projects. Mr. Buehler has broad experience working with downstream refining and chemicals, upstream and LNG. He has conducted energy assessments at Shell America's refinery sites and supported JVs and third-party energy studies. Prior to joining Shell, Mr. Buehler worked for Union Carbide and Dow in operations, the startup of three ethylene crackers, process engineering, feedstock purchasing and JV contract negotiations. He now does consulting work with Petrogenium and Buehler Consulting. He is a member of AIChE and its Ethylene Producers' Committee, and earned a BS degree in chemical engineering from the University of Delaware.

M. BERUTTI, Emerson Automated Solutions, St. Louis, Missouri

Sophisticated analytics build sustainability for renewable diesel projects

Anyone studying the development of the hydrocarbon processing industry over the last century will likely admire how far technologies have advanced, particularly considering the tools that were available in the industry's early years. Many clever people have made discoveries and incremental advances, driving the industry forward to where it is today. Moreover, the larger petrochemical and chemical industries have advanced in parallel, benefiting each other.

While technologies developed during the last decade represent enormous advances from what was available throughout much of the industry's history, one cannot help but wonder if developments might have progressed more quickly had today's tools been available during those formative years.

Spurred by sustainability initiatives in both the public and private sectors, producers around the world are seeking to develop and optimize methods to create diesel fuel without using crude-oil-derived feedstocks. This represents, in effect, a new hydrocarbon industry. However, today's producers, unlike their predecessors, have a full range of modern resources, including digital tools and hardware that were unimaginable just a decade or two ago.

This article will examine what some of these new tools are, and will show how producers are applying them as they develop new processes for manufacturing renewable diesel. Naturally, the lessons learned here can be adopted far more universally, including in established refining and petrochemical processes.

Taming variability. To begin, we should define what process we are talking about (**FIG. 1**), since some common terms are not always clear and are often used interchangeably.

Biodiesel is produced from vegetable oils and animal fats, which are converted to fatty acid methyl esters (FAME) through a transesterification process regulated by ASTM D6751-20a. This process is suitable for large- and small-scale installations, and vehicles can be adapted to run on biodiesel directly. The main drawback of this process is its tendency to gel at low temperatures, thus limiting the amounts that can be added to the conventional diesel distribution chain.

Renewable, or green, diesel is the primary area of interest. It can also be made with oils and fats as feedstocks (and this extends to include crop residues and other biomass) that are treated using a process similar to conventional oil refining. This process produces fuel that is indistinguishable from conventional diesel. Therefore, it falls under the ASTM D975-21 standard specification for diesel fuel, so producers can blend it into the conventional diesel distribution system without limitations.

Sustainable aviation fuel (SAF) is like renewable diesel, and some processes allow operators to make the switch relatively easily when market opportunities become available. Byproducts from the refining process (e.g., light hydrocarbons and naphtha) can be used as non-fossil fuels to meet processing needs. They can be sold as chemical feedstocks or added

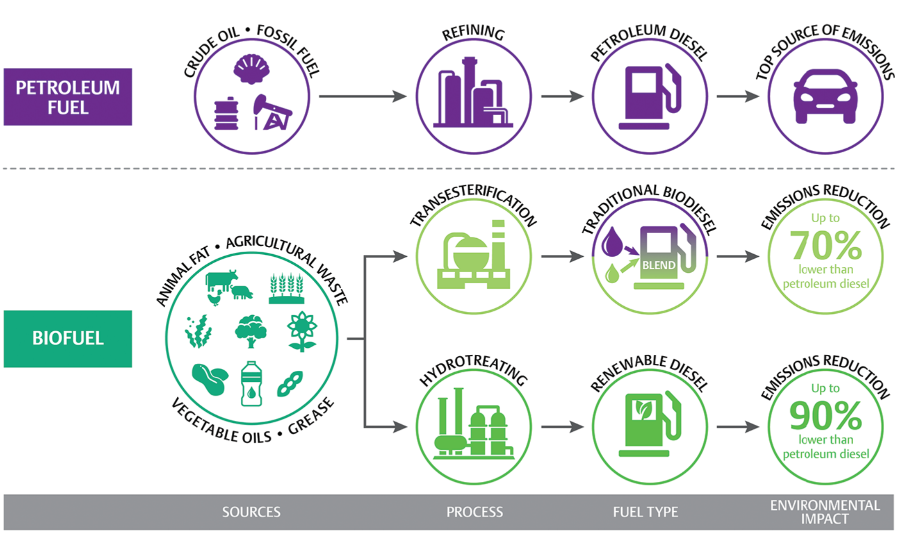


FIG. 1. Alternatives to conventional diesel use the same feedstocks, but with different approaches and results, each with their own characteristics.

to gasoline blending stocks to reduce the operators' carbon intensity score. Production complexity stems from the variety of feedstocks.

To maintain production volumes and continuous operation, refiners must anticipate changing feedstock types. Operators may be working with crop residues one day and animal fats the next, so process parameters can change frequently. This high degree of variability is the primary differentiator from conventional refining and represents the main production challenge.

Data lost or ignored. The following is a deeper dive into today's tools, with three main assumptions:

- Producing renewable diesel is subject to more process variability than conventional refining.
- All the necessary tools are at our disposal to overcome the challenges.
- The same tools can have the same positive effects in many other application areas.

All the elements connected with a renewable diesel production unit produce data in many forms, and it is up to each facility to determine how much data gets collected and retained. Critical process variables will be kept in a historian, but the associated storage costs can drive a company to draw a line between the data it considers useful and the data it does not consider to be useful. The challenge is that many companies underestimate the value of their data because they cannot imagine an impactful way to apply all the data that they have collected.

For the most part, data gets locked in historians that are isolated on operational technology (OT) networks, where the data is stored and processed for release to management in the form of reports. An engineer may occasionally dig into a historian to solve a production problem, but much of the data simply remains in storage, never to be accessed again.

Data lakes bridge communication gaps. Since individuals have access to numerous tools, one may look at many companies and conclude that they do not have enough data, and that they are not putting their data in the right place. To start, historians are not the answer. Historians store data in the format in which it is produced, so data that is not directly comparable stays that way, making analysis more difficult. The solution involves either replacing historians or interconnecting them with a data lake.

The data lake concept is not unique to manufacturing, but using it in OT environments is relatively recent. The same approach is used for the same purpose for corporate information technology (IT) applications; so, to many coming from the IT side, the data lake concept will sound like a logical solution. However, some IT providers have encountered difficulties trying to extend these platforms into OT environments for a variety of reasons. An OT-oriented approach is much more effective and still allows easy extension into IT applications. The ability to connect from OT to IT has generally proven to be easier than the reverse.

How is an OT data lake different than a historian? A data lake aggregates operations data (**FIG. 2**) from manufacturing, process control and IT systems, without disruption, and then contextualizes it and transforms it into actionable information. These results can be presented to every decision-maker on any device, any time and at any location, thus creating manufacturing intelligence.

When data is stored in this type of a central repository, users and critical systems can find anything they need in one place for more efficient analytics, trending and reporting. This eliminates OT data silos, while providing IT personnel with a single location to manage, protect and easily integrate OT data with IT tools and cloud applications.

How does a data lake do these things? A renewable diesel plant or unit has various controllers, valves, process instruments and other equipment types that all individually create their own specific type of data, using their specific format and semantics. Historians can be added to communicate with various devices or classes of devices, but each only communicates with its assigned sources, using its own format and semantics. This leaves data from different sources difficult to compare or integrate, even within a single historian, and multiple historians do not normally communicate to each other. This results in isolated data stored in multiple silos that are cut off from each other.

A data lake takes a different approach. It creates a common data transport layer that supports all OT communications, regardless of the device type. This layer provides a common topology that can bridge gaps, making it far more practical to use digital tools for data analyses.

A modeling studio takes historical data and "cleans" it to remove outliers and create a more correlated data set that is suitable for modeling purposes. Next, it uses machine-learning algorithms to build predictive models. Once these are assembled, it is possible to verify a model's ability to predict anomalies by using test data.

Once validated, the model is deployed in a production environment, using a project studio. This is the area to analyze real-time data and to create rules for complex events. When in operation, if a predictive model finds a new abnormal condition or process anomaly, the model is retrained to consider that condition the next time.

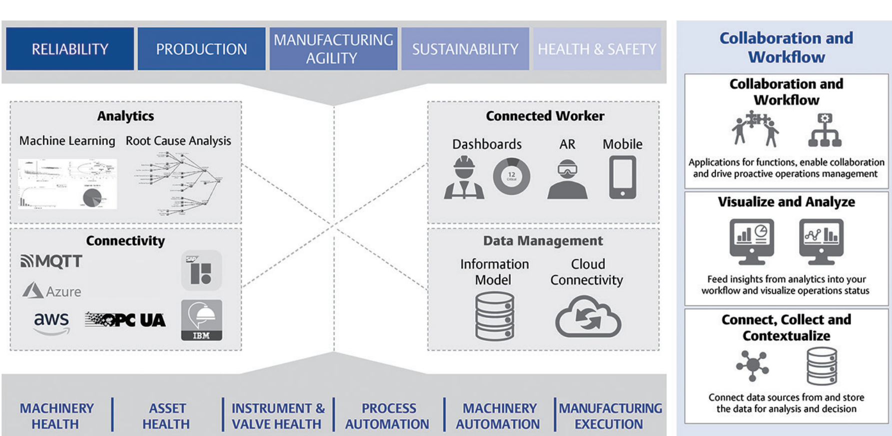


FIG. 2. An analytics platform[a] provides information in a collaborative environment that is able to unify data, people and systems. This can drive improvements in operational performance and workplace efficiency through automated workflows, advanced analytics and enhanced decision support.

This basic approach is the mechanism used to generate information that informs decision-making for actions in areas such as asset health, efficiency and predictive emissions.

What can a data lake facilitate? A data lake is not an end, but rather a means for analytics tools to deliver the desired benefits. A larger implementation, with a data lake and applications, unifies data while connecting systems and people. Together, these drive operational performance and workplace efficiency through automated workflows, advanced analytics and enhanced decision support. This is broad conceptually, so we need to focus on three applications:

- Feedstock variability
- Energy management
- Asset health.

First, renewable diesel production can begin with a wide variety of feedstocks, each of which must be treated differently to convert them into a suitable intermediate. While the variety is greater than with conventional refining processes, different fats or biomass can be separated into categories based on their characteristics. A data lake allows operators to compile a detailed catalog of feedstocks, including process adjustments needed to accommodate each of these feedstocks. Throughout time, analysis of historical process data from daily operations builds predictive models that simplify the steps that operators must take as they facilitate a feedstock change, no matter how drastic a shift this change represents. Once the transition is completed, the predictive model can optimize the process for the duration of that feedstock run.

Second, renewable diesel—like its conventional counterpart—requires energy inputs during manufacturing. The same fired heaters, boilers and the like are used, and these processes are under the same environmental regulatory constraints. Optimizing energy consumption is critically important, calling for detailed analysis of all energy consumers to vary their efficiency. This necessary level of detail depends on extensive instrumentation and data analysis for every installation. One beneficial side effect of having all this data is the ability to use predictive emissions techniques. When sufficient statistical information is available via a data lake, regulatory agencies will often allow a plant to base its emissions monitoring on modeling rather than on actual measurements via analyzers—which is far less expensive than maintaining a fleet of analyzers.

Third, the following will examine operating asset health and a specific centrifugal pump installation (**FIG. 3**). This hypothetical installation has been outfitted with several WirelessHART condition monitoring instruments, including those for measuring bearing temperature, vibration, motor current draw and flow.

Using open communication protocols, the data lake takes operations data from these instruments and then centralizes and contextualizes it, even if the different sources represent different data types. Real-time data is delivered to the maintenance and reliability teams via interactive dashboards so they can see what is happening at any moment. This capability alone would be a major advancement for many plants, but it is only a start. As historical data accumulates, it becomes possible to perform operational analytics.

In our scenario involving this pump, there are two main analytical methodologies—principles-driven and data-driven analytics—and both are available. Each approach has its capabilities and should be matched to the application. In some cases, both can help.

Principles-driven analytics techniques use laws and principles that have been codified for many years. Engineers understand how a pump works (e.g., pump head, speed, pressure differential), and, for example, how the current draw vs. liquid moved provides basic efficiency.

However, the concept can be extended to examine other variables (**FIG. 4**), such as vibration or changes in suction and discharge pressure, both of which can point to abnormal situations indicative of developing problems. With enough data, it is possible to characterize the deterioration by using failure mode and effects analysis. For example, it will be possible to know that when vibration reaches a specific threshold, the installation is nearing failure. This will be based on the specific unit, but will also draw on experiences with other units like it within the plant and perhaps within the larger company.

Simultaneously, data-driven analytics create statistical models to fit the data. For a pump, analytics can create a graph plotting probability of failure on demand (PFD) against operating hours for a given motor/pump configuration. The larger the number of comparable installations, the more reliable the analysis will be. If the company

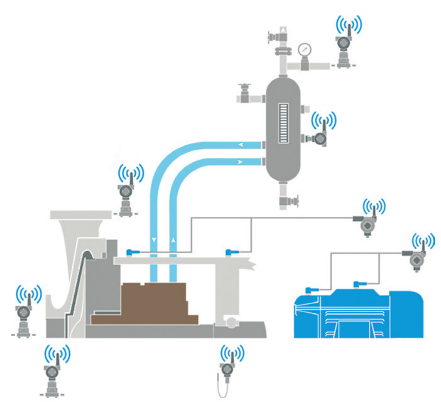


FIG. 3. The centrifugal pump in this example is outfitted with instruments to monitor critical condition indicators.

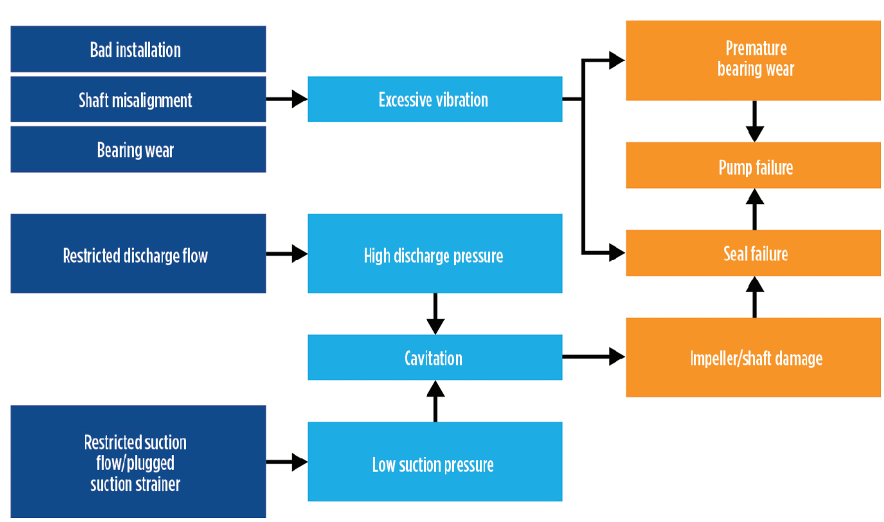


FIG. 4. Many equipment condition monitoring techniques use principles-driven analytics to determine when problems are developing.

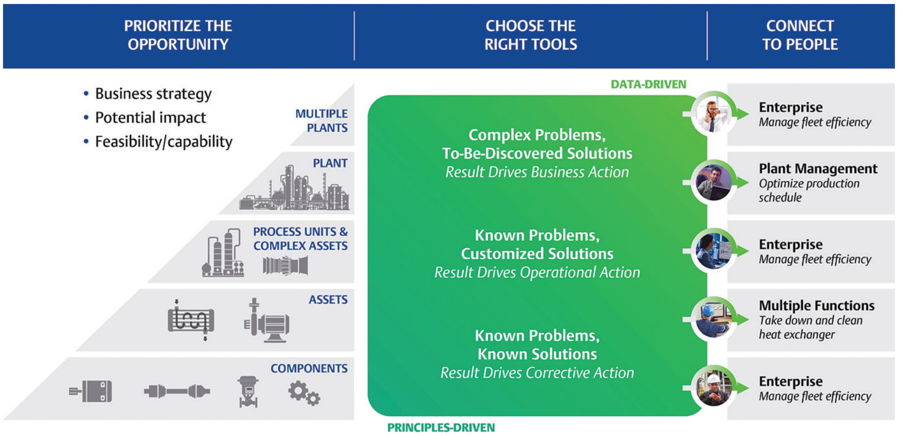


FIG. 5. The ability of an analytics platform to work with a data lake's resources allows for complex problem solving and decision support.

assigns a PFD value based on criticality of a given installation, it will be a simple matter to determine the remaining useful life of an asset, thus allowing the maintenance team to optimize its activities.

While this pump example is real and valuable, the concepts that it illustrates go beyond this limited application and can be extended universally. We can move from this simple case to a much higher level of complexity, effectively treating an entire plant or production unit as a single asset.

The bigger picture. Using analytics to characterize a component—such as our pump example—is the bottom of the pyramid (**FIG. 5**). There are many other examples, but these are the least complex, and they offer, individually at least, the lowest payoff. However, when combined, they have a major financial benefit.

Moving up the pyramid includes more complex and costly assets (such as large compressors, heat exchangers and fired heaters). Beyond these assets are a complete production unit, a plant and, ultimately, a company with multiple plants. As complex as these assets can become, they still follow the same basic methodology of applying data-driven and principles-driven tools.

FIG. 5 illustrates how this works. The left section defines the levels of challenges and opportunities based on their scale and complexity. Moving from the bottom up, the number of items analyzed becomes smaller (a single plant vs. thousands of components), but the complexity and stakes increase. The upper sections are where the importance of analytics supported by data lake technology becomes the most evident. Traditional methods using historians may be sufficient for some components and assets, but the ability to perform useful analysis begins to break down due to data incompatibilities.

The data lake's ability to centralize and contextualize data makes it possible to apply analytical tools to more complex issues, as illustrated in the central green section of **FIG. 5.** As a practical matter, moving higher up the pyramid calls for more data-driven analysis due to complexity, whereas principles-driven methodology is more frequently used at the bottom. Solving higher-level problems often assumes that lower-level problems are resolved, so a bottom-up approach may be preferred. Additionally, addressing lower-level problems can sometimes resolve higher-level issues.

Identifying opportunity. The rightmost column in **FIG. 5** illustrates the kinds of opportunities available using this methodology. Again, moving from the bottom to the top, one starts with individual pieces of equipment, and then gradually works toward enterprise-wide solutions. Management must look at this range and determine where to work on improvements to deliver the greatest impact.

Take efficiency, for example. Unfortunately, no process unit or plant comes equipped with an efficiency knob that operators can easily turn up through a single action. Improving overall efficiency is a process of making dozens and possibly hundreds of small changes that tighten up the operation to improve feedstock utilization, reaction rates and heat recovery, while also reducing losses, among other parameters.

Improving overall efficiency works in a data lake environment because this makes it possible to recognize the opportunity at the upper levels and then to drill down to individual reactors, heat exchangers, boilers, distillation towers and other assets to recognize where inefficiencies exist and how they can be fixed. The required actions happen at all levels and are coordinated by the overall analytics platform.

The unique challenge of renewable diesel production. The aforementioned points apply to companies producing renewable diesel, and to virtually every process industry. It is true that renewable diesel production faces particularly severe challenges related to feedstock variability, but those challenges are increasingly common across many industries.

The challenges for renewable diesel are not entirely technical, as environmental sustainability is also a key component. Often, conventional diesel is easier to make and cheaper to produce. A benefit of renewable diesel is that it is not made from fossil feedstocks and, therefore, it helps to reduce carbon emissions.

Maintaining production efficiency despite major feedstock variability requires the right tools to handle refining complexity. These new OT tools, with a data lake and applications, unify data to produce meaningful insights while connecting systems and people. Together, these tools drive operational performance and workplace efficiency through automated workflows, advanced analytics and enhanced decision support. The challenges are significant, but today's technologies make it is easier than ever to design and implement optimized processes and work practices. HP

NOTE

[a] Emerson PlantWeb Optics platform

MART BERUTTI is the Vice President of Digital Transformation at Emerson Automation Solutions. He has 30 yr of experience in the fields of dynamic process simulation, control system validation, industrial ethernet integration and operator training. At Emerson, before accepting his current role, Mr. Berutti served as Vice President of process simulation. Prior to joining Emerson, he was the President and Chief Operating Officer of MYNAH Technologies from its founding in 2002 until Emerson acquired the company in May 2017. Mr. Berutti earned a BS degree in chemical engineering from the Missouri University of Science and Technology.

K. DARU, Air Products and Chemicals Inc., Houston, Texas; **A. PANCHAL,** Independent Consultant, Mumbai, India; and **P. GALA,** Reliance Industries Ltd., Mumbai, India

Uncommon lessons: Shell-and-tube heat exchangers—Part 2

Shell-and-tube heat exchangers (STHEs) are the most common heat transfer process equipment in all types of industrial plants. Due to the variety of available construction options, an STHE can be suitably designed for most processes and design conditions and can be constructed from different materials or combination of materials.

Part 1 of this article (published in the April 2022 issue) began the focus on specific types of lessons and design limitations that none of the commercial design programs (software, Excel spreadsheets, etc.) will identify with their "help" menus or warnings. Merely checking drawings with respect to design calculations is insufficient to ensure thorough compliance and reliable product design that will operate (thermally and mechanically) as intended.

The authors' purpose is to shed light on those uncommon but valuable lessons that are not discussed in reference manuals or design guides. These lessons will be helpful in engineering and design activities.

Sleeve welded to the inner torus of an expansion bellow.

An expansion bellow is a flexible element and the knuckled (torus) section of the bellow is a highly stressed region. The torus sections of the expansion bellow are responsible for the flexible behavior of the joint in service.

The welding of clips (like the liner or sleeve) on the torus section should be avoided because it increases the effective stiffness of the expansion bellow. With stiffeners attached to the knuckled sections, the joint requires more bending force or longitudinal movement to flex the bellow—this may not exist based on the estimated differential thermal expansion—and this can result in overstressing the tube-to-tubesheet joint.

In earlier ASME Code[1] editions (before 2015), no explicit guidance was available on whether it was possible to weld sleeves on the torus regions, or, if it was possible, what the maximum sleeve thickness was that could be welded. A query was submitted to the code, and it now has incorporated guidance on the maximum thickness of the sleeve that may be welded on the annular portions of the bellows.

Exchangers constructed prior to 2015 were provided with internal sleeve construction based on the vendor's engineering experience and/or judgement, as no guidelines were available in the applicable ASME Code editions.

The authors encountered sleeve constructions (as shown in **FIG. 5A**) for an exchanger designed and constructed per the 2015 edition of the ASME Code[1] in which no requirements for sleeve provisions were specified. As the thickness of the sleeve was not limited by Code, sleeves were provided without considering the impact of welding these sleeves on the performance of the bellows.

In the 2017 edition of the ASME Code, a paragraph was introduced that permitted welding the sleeve to the bellow, if required, with specific thickness limitations. In many cases, the thickness limitation per the ASME Code may be difficult to meet for practical reasons. For such situations, an alternate arrangement is used by introducing a transition piece that eliminates welding of the sleeve to the expansion joint, as shown in **FIG. 5B.**

Impingement rods.

As per API Standard 660[3], impingement rods should be supported on both ends and not attached to the tubesheet by welding alone. For proper strength, the impingement rods must be primarily tapped inside the

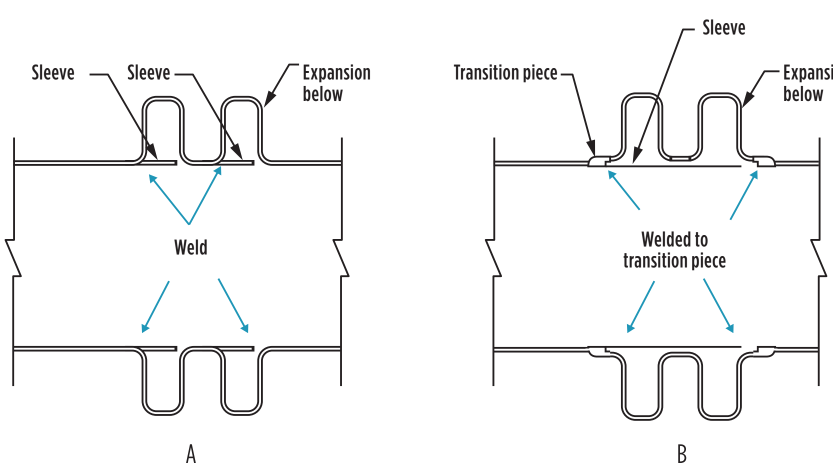


FIG. 5. Arrangement of a sleeve for an expansion bellow.

tubesheet thickness—this is more rigid compared to tack welding impingement rods on the back face of the tubesheet. The authors have encountered heat exchangers in which impingement rods were fillet- or tack-welded; due to a lack of accessibility, the welder may have been unable to deposit a continuous fillet weld. This quality deficiency caused the impingement rods to get dislodged/separated from their attachment welds during service. The separation of impingement rods resulted in oscillating deflection of the rods, which kept striking the tubes underneath, ultimately causing failure of the tubes.

The operating company using this type of exchanger in the plant detected tube leakage that led to the contamination of the shell-side fluid due to leakage of tube-side fluid into the shell-side stream. The exchangers required an urgent offline repair that forced an unscheduled shutdown.

The API standard provides narrative guidance on how the impingement rods should be installed. The arrangements shown in **FIG. 6** may be interpreted as meeting API narrative guidelines.

Bundle retainer not provided for kettle-type heat exchangers. TEMA[2] Paragraph RCB-4.9 provides non-mandatory requirements for a bundle retaining arrangement and permits other arrangements that serve a similar purpose. The authors observed that the bundle retaining or equivalent device was not shown in the detailed fabrication drawings of some kettle-type heat exchangers.

It was also determined that the requirements of the bundle retainer were not specified in the fabricator's scope by the purchase order. Excluding the bundle retainer on such types of exchangers may cause shifting (dislocation) of the bundle from its correct position during transportation or unexpected plant operation conditions.

The bundle retainer is an essential device for alignment and hold-down purposes of the bundle for kettle-type exchangers. According to TEMA[2], it is recommended that a bundle retainer or equivalent device be provided at the last baffle over the u-bend. The retainer helps the alignment of the bundle along the shell and acts as a hold-down device for the bundle because there is significant room for bundle misalignment or movement in the kettle-type shell.

Using a Class 2 forged tubesheet with hub for low-alloy steel exchangers. This unique design lesson could have significant cost and schedule impacts if it is uncovered during fabrication. In a few exchangers from low-alloy steel materials, the authors noticed the hub of the low-alloy steel (LAS) forged tubesheet was given the same thickness as the adjoining low-alloy steel cylindrical shell to which it was butt-welded.

This issue is particularly found in LAS heat exchangers where the shell material

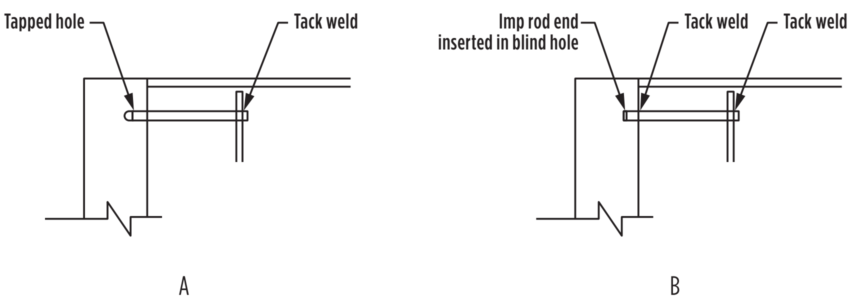


FIG. 6. Fixing arrangement for an impingement rod.

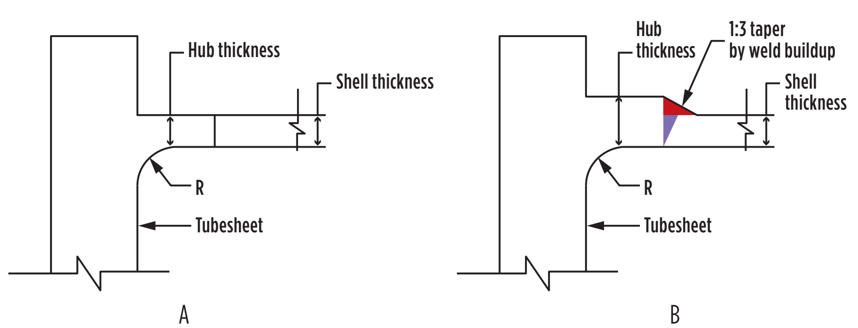


FIG. 7. Hub radius detail.

TABLE 3. Section III, Classes 2 and 3;* Section VIII, Div. 1 and 2;** and Section XII maximum allowable stress values (*S*) for bolting materials

Line number	Nominal composition	Product form	Spec. number	Type/grade	Alloy designation/ UNS number	Class/condition/ temperature	Size/thickness, mm
	Ferrous materials						
1	Carbon steel	Bolting	SA-307	A	–	–	$6 \le t \le 100$
2	Carbon steel	Bolting	–	B	–	–	–
3	Carbon steel	Bolting	SA-449	1	K04200	–	$38 < t \le 75$
4	Carbon steel	Bolting	SA-325	–	–	–	–
5	Carbon steel	Bolting	SA-325	1	K02706	–	$32 < t \le 38$
6	Carbon steel	Bolting	SA-449	1	K04200	–	$25 < t \le 38$
7	Carbon steel	Bolting	SA-354	BC	K04100	–	$64 < t \le 100$

*See maximum temperature limits for restrictions on class **Use with Part 4.16 of Section VIII, Div. 2

is SA-387-11-2 plate and the tubesheet is integrally welded to the shell. The corresponding tubesheet forging normally adopted is from SA-336-11-2, and the tubesheet is provided with a hub for welding with the shell from SA-387-11-2. The tubesheet hub thickness was inadvertently maintained the same as the shell thickness without realizing that the allowable stresses of LAS Class 2 forging are lower than that of LAS Class 2 plate. For such constructions, the tubesheet forging should be from SA-336-11-3 (Class 3), which will match the allowable stresses of the LAS Class 2 plate; the hub can be from the same thickness as the shell. **FIG. 7A** shows the tubesheet hub-to-shell attachment detail when both thicknesses are equal (the hub material allowable stress is equal to the shell material allowable stress), while **FIG. 7B** shows the attachment detail when the hub (material with a lower allowable stress) is welded to the shell (material with higher allowable stress compared to the hub).

This will be difficult to correct (or weld repair) once the tubesheet hub is machined out and it is determined that the Class 2 forged hub thickness is maintained the same as the cylindrical shell thickness from a Class 2 plate.

Limitations about pressure vessel bolting diameters. Pressure bolting materials for ASME pressure vessel applications are limited by the diameters of bolt material specifications listed in the "Allowable Stress Table 3" of Section II-Part D. Whenever bolt specifications from ASME Section II-Part A or Part B are used for pressure vessel application, only those diameters of the material specifications that are listed in Table 3 of Section II-Part D[5] should be used.

The authors encountered a heat exchanger in which the body flange used a 2.75-in. diameter SA-320-L7 as pressure bolting. ASME Code Section II-A may list mechanical properties and chemical composition of a bolt specification for a diameter that may be outside the size range listed in ASME Section II-D Table 3; this does not mean that such a bolt diameter and specification are permitted for use in pressure vessel applications.

Even if ASME Code Section II-A permits a 2.75-in. diameter SA-320-L7 bolt for manufacture, allowable stresses for SA-320-L7 bolts are only provided up to 2.5-in. diameter maximum in the Allowable Stress Table 3 of Section II-D[4]. Therefore, 2.75-in. diameter SA-320-L7 bolts are not permitted for pressure vessel applications.

For any bolt material, the size/thickness column should be used as a reference for diameter limitations of bolts. Refer to the extreme right column in **TABLE 3,** an extract from Table 3 of ASME Section II-D[4].

SA-456-660 bolts are required to be impact tested at minimum design metal temperature (MDMT) of ≤ 20°C. SA-456-660 A or B bolts are used as pressure bolting on high-alloy body flange joints at high temperatures [> 650°F (> 343°C)], particularly under high design pressure (> 50 barg). The advantage of Alloy 660 is that there is no diameter restriction in Section II-D, and the allowable stresses are steady up to

1,000°F (538°C) for Div. 1 applications. These bolts are also used on high-alloy floating heads as corrosion resistant alloy bolts because they can remain fully submerged in—or exposed to—shell-side corrosive fluid. Most other high-alloy bolting material options have low allowable stresses and/or diameter limitations for pressure vessel applications.

SA-453 is a precipitation-hardened austenitic stainless steel. The precipitation hardening process involves the strengthening of the high alloy by specific thermal treatment. Since these bolts are thermally heat treated, they are required to be impact-tested quality if the MDMT is ≤ 20°C per UHA-51 of ASME Code Section VIII-1. While the impact testing requirement is clearly specified for SA-453-660 bolts in Div. 2, it is not so explicitly stated in Div. 1.

The authors noticed SA-453 bolting used for heat exchanger body flanges without being specified for impact testing. Considering these exchangers were designed under Div.1, the impact testing requirement was missed. The authors believe that the manufacturer considered SA-453 bolts as austenitic stainless alloy bolts and interpreted that they are exempt from impact testing under the same paragraph UHA-51 of Section VIII-1. The code paragraphs would be helpful to identify impact testing requirements, as applicable, if SA-453 are viewed as "thermally heat treated" austenitic stainless steels, which they are.

The following response was received from the code committee on the authors' inquiry #15-2817:

> Standard designation:
> BPV Section VIII Div. 1
> Edition/addenda: 2013 Ed.
> Paragraph/figure/table number:
> UHA-51(c)
> Subject description:
> Section VIII, Div. 1; UHA-51(c)—
> Thermal treatments for material
> Date issued: 05/26/2016
> Record number: 15-2817
> Interpretation number:
> BPV VIII-1-16-45
>
> Question(s) and Reply(ies):
> **Question:** Do the thermal treatments listed in UHA-51(c) include thermal treatments performed by the material supplier during the manufacture of the material in addition to thermal treatments performed during the manufacture of the vessel?
>
> **Reply:** Yes

Slot length in sliding saddle insufficient for possible thermal expansion.

Horizontal vessels and heat exchangers are provided with saddle supports per engineering specifications, which are often part of applicable purchase order documents. If the supported vessel is abnormally long and/or the exchanger operates at elevated temperature and has low environmental MDMT, the slot length indicated for the sliding saddle in the engineering specification may be insufficient for the full range of thermal expansion.

The authors observed exchangers in which the slot length provision on the sliding saddle was insufficient for the anticipated thermal movement of the sliding saddle based on the operating temperature range. The engineering specification was used to specify the slot length in the sliding saddle base plate, and it was not verified if the standard slot length would be adequate to cover full thermal movement. An insufficient slot for the sliding saddle can lead to its restricted movement during thermal growth of the shell, for which the exchanger saddle/shell is not designed.

Takeaway. Numerous design reviews of equipment data were performed in the authors' routine engineering tasks; often, such reviews are performed hastily. A pronounced shortage of quality time for reviews always exists because the documents are pushed out promptly under the pretext that checking for design adequacy is someone else's responsibility, or not enough hours are allocated for thorough review.

On the other end, an equipment design that meets the code as well as relevant and accepted good engineering practices (RAGAGEPs) is the goal for equipment supply so it will perform as expected throughout its design life, and even beyond. A thorough independent review plays a key role in achieving this tall order; a cursory review will not help achieve this goal. The objective for engineers and reviewers should be to leave no stone unturned during reviews to achieve the required result, regardless of the time it takes to provide the optimum engineering services—there is no substitute for quality. Proper design review experience is acquired by continuous practice and improvement, and the authors feel that the feedback of uncommon lessons will be helpful.

LITERATURE CITED

[1] American Society of Mechanical Engineers (ASME) Boiler and Pressure Vessel Code, Section VIII, "Rules for construction of pressure vessels—Div. 1," New York, New York, July 2019.

[2] "Standards of the Tubular Exchangers Manufacturer's Association (TEMA)," 10th Ed., 2019.

[3] American Petroleum Institute (API) Standard 660, "Shell-and-tube heat exchangers," 9th Ed., Washington, D.C., May 2020.

[4] American Society of Mechanical Engineers (ASME) Boiler and Pressure Vessel Code, Section II-D, "Material properties," New York, New York, July 2019.

[5] American Society of Mechanical Engineers (ASME), Boiler and Pressure Vessel Code, B31.3, "Process piping," New York, New York, July 2018.

[6] American Society of Mechanical Engineers (ASME) Boiler and Pressure Vessel Code, Section II-A, "Ferrous material specifications," New York, New York, July 2019.

KUNTAK DARU is a Pressure Vessel Engineer with Air Products and Chemicals Inc. Mr. Daru works in the mechanical equipment division of Air Products, supporting engineering and design activities of AP projects worldwide from the Houston, Texas location. Prior to working with Air Products, Mr. Daru was a Technical Director in mechanical engineering at Fluor Corp. in Sugar Land, Texas. He has more than 30 yr of experience in the engineering, design and specification of static equipment in the refining, oil and gas, petrochemical and chemical industries. Mr. Daru has produced more than 10 publications in his field and holds a BS degree in mechanical engineering from SVNIT in Surat, India. Mr. Daru is a Registered Professional Engineer in the states of Texas and Louisiana. The author can be reached at Kuntak.Daru@gmail.com.

ANILKUMAR PANCHAL is an Independent Consultant in Mumbai, India. He has more than 15 yr of hands-on experience as a Mechanical Design Engineer in butyl rubber, gasification, ethane cracker, chlorinated PVC and various revamp projects. He previously worked for L&T Heavy Engineering, Bechtel India Ltd., Jacobs Engineering, Reliance Industries Ltd. and KIPIC, Kuwait. He holds a BS degree in mechanical engineering and an MS degree in industrial process equipment design. He is a Certified Chartered Engineer and Professional Engineer in India. The author can be reached at anil0208@gmail.com.

PUNITA GALA is a Senior Design Engineer at Reliance Industries Ltd. and is part of the project management team. She has worked on benzene recovery units, extract hydrotreater units, SCAN-Finer units, as well as pressure vessels, shell-and-tube heat exchangers and packaged items. Prior to Reliance Industries Ltd., she worked for Bechtel India and Larsen & Toubro Ltd. Ms. Gala graduated from L. D. College of Engineering, Ahmedabad, India, in mechanical engineering.


Technology and Business Information for the Global Gas Processing Industry
GAS PROCESSING
& LNG
GasProcessingNews.com | MAY/JUNE 2022

Lee Nichols,
Editor-in-Chief/
Associate Publisher

Modern society would not be possible without the use of pipelines to transport natural gas, crude oil and finished products to demand centers. These products provide much-needed feedstocks for processing operations, as well as fuels for transportation, heating and electricity.

The need for additional pipeline systems is imperative, especially with growing economies in many different parts of the world—demand for heating, electricity and cooking are leading to a significantly higher demand for natural gas.

To satisfy future demand, new natural gas pipelines systems are being built to transport this commodity from import centers or domestic production areas to demand centers. Presently, Gulf Energy Information's Global Energy Infrastructure database is tracking more than 430 active natural gas pipeline projects around the world. These systems represent tens of thousands of miles of pipelines and constitute hundreds of billions of dollars in capital investments.

The increased use of natural gas is also providing countries around the world with new options to decarbonize various industries. Many nations are actively pursuing low-carbon initiatives to reach future low- and net-zero economies. Increasing natural gas in a country's total energy mix only helps on this pathway to lower carbon emissions.

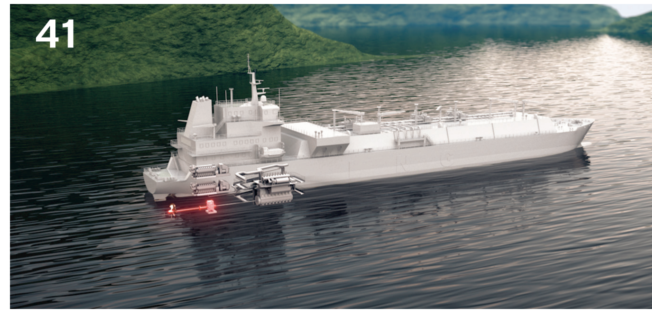
Since the use of natural gas pipelines is a crucial part of modern societies, *Gas Processing & LNG* has devoted this issue's Special Focus section to new technologies that can enhance pipeline operations. **GP**

GAS PROCESSING & LNG
www.GasProcessingNews.com

P. O. Box 2608
Houston, Texas 77252-2608, USA
Phone: +1 (713) 529-4301
Fax: +1 (713) 520-4433
Editorial@GasProcessingNews.com

CONTENTS

GasProcessingNews.com | MAY/JUNE 2022



Cover Image: Gori-Kareli gas pipeline. Photo courtesy of Georgian Oil & Gas Corporation.

Wärtsilä awarded EPC contract for Antigua LNG terminal

Antigua Power Co. Ltd. has awarded Wärtsilä an EPC contract to build a new LNG terminal on the island nation of Antigua. The project—being developed by U.S.-based Eagle LNG—combines a 46-MW dual-fuel power plant and an LNG terminal, storage and regasification facility. The plant will operate with five Wärtsilä 34DF dual-fuel engines capable of operating with both gas and light fuel oil. The new power plant will help satisfy increasing power demand on the island. The plant is expected to begin operations in 3Q 2023.

Tellurian begins construction on Driftwood LNG project

In late March, Tellurian gave the greenlight to EPC firm Bechtel to begin limited construction on the 27.6-MMtpy Driftwood LNG project. Located near Lake Charles, Louisiana (U.S.), the multibillion-dollar project will export U.S. natural gas to international destinations. Current Phase 1 construction includes site preparation and laying foundations. Once financing is secured, Tellurian plans to moves forward with additional Phase 1 activities, including completing the construction of two LNG trains with a total processing capacity of 11 MMtpy. The operator plans to begin LNG exports in 2026.

NextDecade aims at FID for Rio Grande LNG project in 2H 2022

In March, NextDecade has signed a Heads of Agreement (HoA) contract with Chinese energy provider Guangdong Energy. The energy company agreed to purchase 1.5 MMtpy of LNG from NextDecade's Rio Grande LNG terminal. The HoA contract brings NextDecade one step closer to reaching FID on the project, which the company intends to do in the second half of this year. If greenlighted, Train 1 operations are expected to commence in 2026.

Woodfibre LNG to commence construction

Woodfibre LNG plans to begin construction on the $1.6-B LNG export terminal in 2023. The 2.1-MMtpy facility will be located southwest of Squamish, British Columbia, Canada and will be built by McDermott—the company was awarded an engineering, procurement, fabrication and construction contract in late 2021. The terminal is scheduled to start commissioning in 2027.

Italy to possibly use two FSRUs to cut Russian dependence

To help wean off Russian natural gas supplies, Italy may install two floating storage and regasification units (FSRUs). The two vessels, with a total capacity of 10 Bm^3y, will be in the Tyrrhenian Sea and the Adriatic Sea. At the time of this publication, a final investment decision has not been made on these projects.

Italian oil and gas company, Eni, has also announced a deal to source additional natural gas supplies from the Congo Republic. Eni plans to import an additional 4.5 Bm^3y of natural gas from the African nation and will ramp up natural gas production within the Congo Republic—the company is part of an LNG project that is set to begin operations in 2023.

To mitigate Russian natural gas supplies, Italy has also signed new natural gas imports deals with Angola, Algeria and Egypt.

South Africa to build LNG terminal in Richards Bay

Transnet National Ports Authority (TNPA) has announced that a tender to build a new LNG import terminal in Richards Bay, South Africa has been sent out to potential companies around the world. Once it is operational in 2026, the terminal will help import additional natural gas feedstock to help mitigate electricity shortages within the country.

Germany to lease up to four FSRUs to satisfy domestic natural gas demand

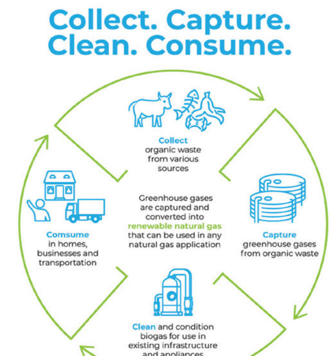
To help wean off natural gas imports from Russia, Germany has announced plans to lease up to four FSRUs. These floating vessels will aid the country in importing natural gas until new terminals are built. At the time of this publication, three of the FSRUs are scheduled to begin operations by early 2023.

Germany has also announced plans to invest more than $3 B to build four LNG import terminals over the next decade.

Cheniere begins Stage 3 work on Corpus Christi LNG terminal

Cheniere has awarded Bechtel a lump-sum EPC contract for Stage 3 operations at the company's Corpus Christi Liquefaction terminal in Corpus Christi, Texas (U.S.). Stage 3 operations include the construction of seven mid-scale LNG trains, each with a processing capacity of nearly 1.5 MMtpy. Once built, Stage 3 will increase the terminal's capacity by more than 10 MMtpy, reaching a total installed capacity of approximately 25 MMtpy. The expansion project is scheduled start LNG shipments in 2025.

TC Energy and GreenGasUSA to collaborate to build RNG transportation hubs

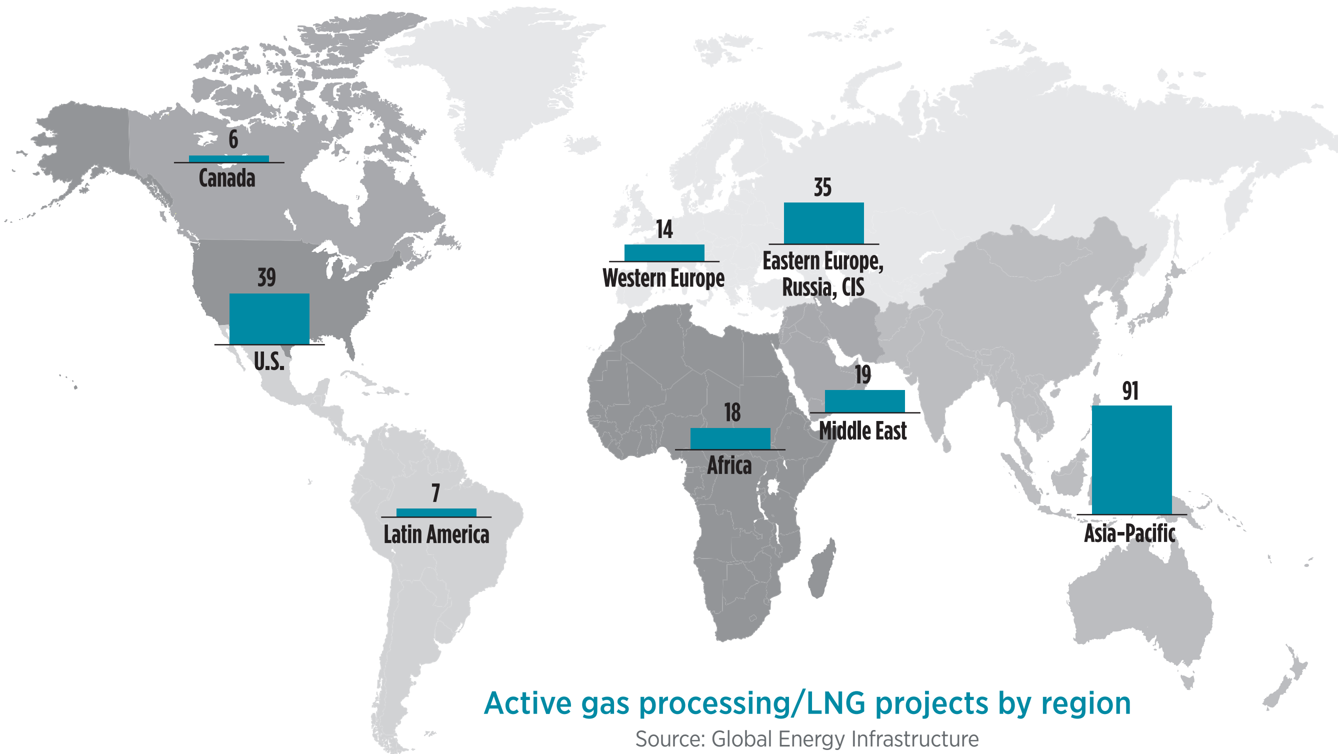


TC Energy and GreenGasUSA have partnered to explore the development of a network of natural gas transportation hubs. These transportation hubs will provide centralized access to existing energy transportation infrastructure for renewable natural gas (RNG) sources, wastewater treatment facilities and landfills. RNG is a clean and reliable fuel that is often made from animal waste. According to TC Energy, as animal waste breaks down, it produces methane. The methane is captured—eliminating it from the atmosphere—and then processed to use as an energy source.

According to the JV's announcement, GreenGasUSA will process RNG to pipeline quality and transport it to the RNG hub. TC Energy will build, own and operate the RNG transportation hubs. Combining TC Energy's vast energy distribution network with GreenGasUSA's extensive experience in gas compression, compressed natural gas (CNG) transportation and pipeline injection is expected to further increase methane capture and the use of renewable fuels within the energy mix. These transportation hubs are expected to be under development in several states along TC Energy's 32,700-mi U.S. natural gas pipeline system within the next 4 yr. The first hub is targeted for service in 2Q 2023.

LEE NICHOLS, Editor-in-Chief/Associate Publisher

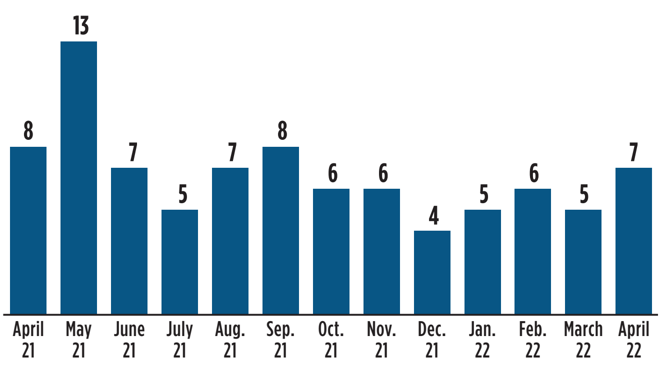
Gulf Energy Information's Global Energy Infrastructure (GEI) database is tracking nearly 230 active gas processing/LNG projects around the world. At 40%, the Asia-Pacific region continues to be the leader in active gas processing/LNG project market share, followed by the U.S. (17%). Over the past year, the GEI database has recorded nearly 90 new gas processing/LNG projects around the world. Most of these projects are within the Asia-Pacific region, which is investing heavily in new gas processing infrastructure to reduce emissions for power generation. Over the past several months, increases in new LNG terminal investments have been announced in Western Europe, as the region is looking for pathways to mitigate reliance on Russian natural gas supplies. **GP**



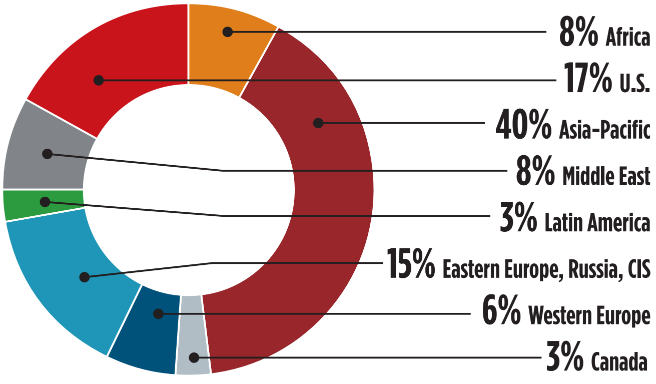
Active gas processing/LNG projects by region

Source: Global Energy Infrastructure

New gas processing/LNG project announcements, April 2021–April 2022



Active gas processing/LNG project market share by region



Engineering assessments in support of pipeline safety: Emerging trends and approaches—Part 1

B. MITTELSTADT, Dynamic Risk, Houston, Texas;
and **D. WILLIAMS,** Dynamic Risk, Calgary, Alberta, Canada

Engineering assessments will increasingly be applied in the pipeline industry to demonstrate related fitness-for-service and appropriate safety margins to stakeholders, including industry, regulatory agencies and the public.

From a U.S. perspective, driving factors include the release of Part 1 of the Pipeline and Hazardous Materials Safety Administration's (PHMSA's) Gas Mega Rule[a] in October of 2019 (effective July 1, 2020), along with pending rulemaking on class location designation changes (this proposed rule would add an alternative set of requirements that operators could use, based on implementing integrity management principles and pipe eligibility criteria, to manage certain pipeline segments where the class location has changed from a Class 1 location to a Class 3 location); and Part 2 of the Gas Mega Rule, which focuses on repair requirements for high-consequence areas (HCAs) and non-HCAs. In lieu of replacing or re-testing pipelines, operators have the option of applying engineering critical assessments and advanced fitness-for-service solutions for validating and reconfirming maximum allowable operating pressure (MAOP), supporting special permitting for class location changes, and varying from prescriptive response and repair requirements.

From a Canadian perspective, operators will continue to be guided by the Canadian Standards Association (CSA) Z662 requirements to apply engineering assessments for demonstrating safe pipeline operations. The application of risk acceptance criteria, such as those proposed for the CSA Z662:23 standard[1] or those considered in PHMSA's "Paper study on risk tolerance,"[2] will support a strategic framework for improving pipeline safety and the use of risk assessment results (i.e., in comparison to acceptable risk thresholds) within an engineering assessment.

Compliance is only part of the picture, as changes in corporate governance and public expectations demand improved safety and reliability. Without advanced engineering assessments, compliance efforts may lead to the replacement of pipe that is otherwise fit-for-service, resulting in missed opportunities to truly improve pipeline reliability and safety.

This article presents an overview of driving factors and trends for engineering assessments and code-related considerations. Practical data integration, approaches and application strategies, as well as a framework for developing and applying effective engineering assessments are provided. Approaches for integrating the data collected from continuously improving inline inspection (ILI) technologies into the engineering assessments are demonstrated.

Key factors. The increased application of robust engineering assessments and fitness-for-service evaluations is expected to play out in several key ways, such as:

- **Increasing pipeline safety and reliability performance:** This boils down to achieving the next level of safety through focused engineering work, along with the development and application of technology, to enhance data gathering and integration, assessment techniques and the associated workflows.
- **Addressing the new and emerging regulatory landscape designed to help achieve improved safety and reliability performance:** Without the insight and analyses from engineering assessments, compliance may result in pipe replacement, pressure reductions and pressure testing (and other material actions) on pipe that is otherwise fit-for-service.
- **Engineering assessments provide an effective performance-based pathway forward:** Detailed engineering and fitness-for-service evaluations will provide the best pathways forward to obtain the greatest yield from resources in terms of increasing pipeline safety and reliability and providing the biggest return on investment.

This article is focused on the following key topic areas as they pertain to engineering assessments and engineering critical assessments:

- Driving factors and trends
- Code-related considerations
- Data integration
- Framework strategies
- Application.

Background. It is important to start out by reviewing definitions related to engineering assessments and engineering critical assessments. There are also some interesting differences worth mentioning in how these are considered in the CSA standard vs. in the U.S. Code of Federal Regulations (CFR).

The engineering assessment and the engineering critical assessment are both defined in the CSA Z662 standard. The CSA Z662 definitions distinguish engineering assessments and engineering critical assessments from each other:

- Engineering assessment: *A documented assessment of the effect of relevant variables upon fitness for service or integrity of a pipeline system, using engineering principles, conducted by or under the direct supervision of a competent person*

with demonstrated understanding and experience in the application of engineering and risk management principles related to the issues being assessed.[1] In the Canadian standard, an engineering assessment applies to demonstrating fitness-for-service in a variety of circumstances.
- Engineering critical assessment: *An analytical procedure, based on fracture mechanics principles, that allows determination of the maximum tolerable sizes for imperfections in fusion welds.*[1] In the Canadian standard, an engineering critical assessment applies specifically to the consideration of imperfections in girth welds after construction. Guidance on the application of engineering critical assessments to girth welds is provided in Annex J of the standard.

The U.S. codes do not specifically define the term *engineering assessment;* however, *engineering critical assessment* is defined in 49 CFR Part 192.3 (referenced in Subpart L, sections 192.624 and 192.632)[3] as part of the Gas Mega Rule and new MAOP reconfirmation requirements:
- Engineering critical assessment: *A documented analytical procedure based on fracture mechanics principles, relevant material properties (mechanical and fracture resistance properties), operating history, operational environment, in-service degradation, possible failure mechanisms, initial and final defect sizes, and usage of future operating and maintenance procedures to determine the maximum tolerable sizes for imperfections based upon the pipeline segment maximum allowable operating pressure.*[3] In the U.S. code, an engineering critical assessment is not specific to flaws in girth welds and has a broader fitness-for-service intent across various threat and defect types that overlaps with that of engineering assessments in the Canadian standard.

In summary, whether we are talking about the CSA's engineering assessment or the U.S. CFR engineering critical assessment, these are both detailed fitness-for-service assessments that can consider various threats and defect types. This article focuses on these types of assessments, in the context of thorough fitness-for-service assessments specific to the circumstance at hand, and they will generally be referred to as engineering assessments in this article.

The circumstances for, and the use of, engineering assessments (Canada) and engineering critical assessments (U.S.) are converging in Canada and the U.S., including, but not limited to, the following:
- Class location designation changes
- Licensed MOP for Canada, approved MAOP for U.S. or MAOP upgrades
- Defect assessments or evaluations of damage
- Flow reversal or change in service
- Reactivating or repurposing a pipeline
- Pressure testing an existing pipeline
- Valve location and spacing
- MAOP validation and reconfirmation (U.S.)
- Special permits (U.S.)
- Governance (safety and reliability case management)
- Return to service after failure or outage.

Class location designation changes are expected to continue to be a key driver for engineering assessments and engineering critical assessments in both Canada and the U.S., respectively. The industry is spending hundreds of millions of dollars every year on pipe replacements where the existing pipe does not align with design requirements for the new class location designation.

CSA Z662 Clause 10.7.2 requires a conformance check of five key elements (mainly design related) for pipe in a class location designation change area. This includes the design factor or location factor, valve spacing, depth of cover, pressure testing and evaluation, and repair of imperfections. While provisions exist for engineering assessments to be carried out if one or more of the five elements cannot be met, it is acknowledged that deviations from design-based elements can create a challenge for gaining stakeholder approval and can result in replacement or pressure reduction actions that are not necessarily performance based or that can measurably improve pipeline reliability and safety.

Similarly, in the U.S., current approaches to evaluate a pipeline's fitness-for-service in class location designation change circumstances are very arduous via the special permitting process [or potentially, in the future, through the current Notice of Proposed Rulemaking (NPRM) related to class location change requirements].

Reconfirmation of MAOP has the potential for creating billions of dollars of effort/impact if the industry must rely on replacements, pressure reductions and pressure tests for compliance. Engineering assessment methodology or other technology solutions can provide an alternative pathway forward when there is sound technical evidence that the pipe is otherwise fit-for-service. There are some cases where pipe replacement is required, but the engineering assessment should support this.

Driving factors and trends. There is a universal mandate from corporate governance structures, new and emerging regulations, and the public to increase pipeline safety and reliability performance. The success of these efforts is critical to maintain the industry's social license to operate. It is important to note that the industry has made great strides in improving pipeline safety and reliability performance through compliance with integrity management, operational and maintenance regulations, and the application of initiatives to move beyond compliance only. Although the industry, regulatory community and public have been diligently working to determine how best to increase pipeline safety performance, all stakeholders acknowledge that the next level of performance will not be easy to accomplish from technical, operational and commercial perspectives.

Without the insight and analyses from engineering assessments in application with improved technology for enhanced data gathering/integration and assessment, compliance may result in pipe replacements, pressure reductions, pressure testing and other material actions on pipe that is otherwise fit-for-service. These massive resource allocations can be otherwise directed more effectively toward mitigating priority safety and reliability risks.

The U.S. regulatory structure continues to go through significant changes that will facilitate gaining the next level of pipeline safety and reliability performance. Part 1 of the Gas Mega Rule contains changes to the regulations for gas transmission pipelines, along with new requirements for the verification of pipeline materials. This part of the rule aims to improve safety with the transportation and operation of onshore gas transmis-

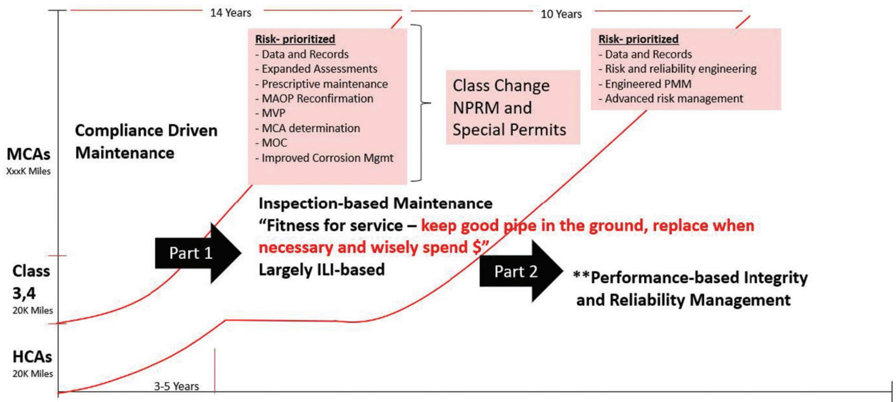


FIG. 1. Pipeline integrity transformation driven by the Gas Mega Rule and NPRM.

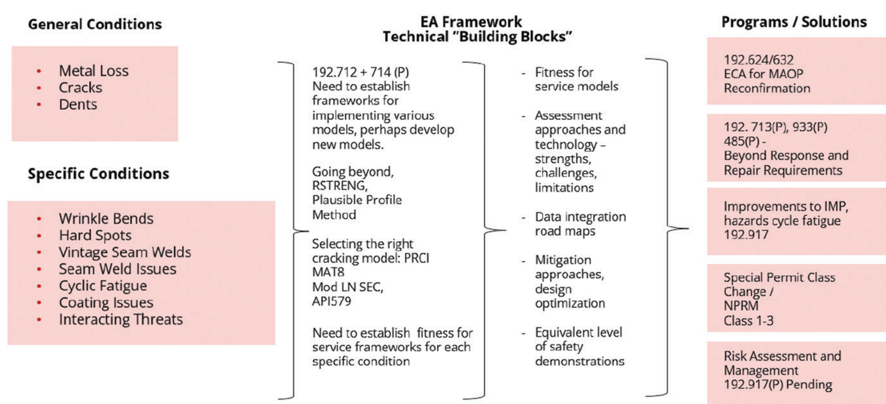


FIG. 2. Engineering assessment technical building blocks for effective solutions.

sion pipelines. Operators will be required to reconfirm the MAOP of their pipelines to ensure their safety, including pipelines that had been exempted from previous safety rulings. Part 2 of the Gas Mega Rule contains new repair requirements for HCAs and non-HCAs. Similar rulemakings have been made or are underway in the U.S. for liquids pipelines, gathering pipelines and gas storage facilities.

Presented as a forward-looking case study, an anticipated generational transformation driven by the Gas Mega Rule and NPRM activities is outlined as a phase diagram in **FIG. 1** for the approximately 300,000 mi of U.S. gas transmission pipelines. Looking at the y-axis, integrity management has largely been focused on approximately 20,000 mi of pipelines in HCAs, with that focus largely based on inspection-based maintenance as illustrated moving along the x-axis. The other 280,000 mi have been managed via design, construction, and operating and maintenance considerations, recognizing that many operators have moved beyond this. Going forward, the HCA miles will begin the journey in transforming from inspection-based maintenance to performance-based integrity management. Pipelines in Class Location Designation 3 or 4, moderate consequence areas, and, most likely, most of the remainder of pipelines will enter inspection-based maintenance and advanced fitness-for-service approaches.

Although it has some limitations, the Gas Mega Rule has provided a strategic framework for improving pipeline safety and the use of engineering assessments, represented by the center portion of the phase diagram.

Detailed engineering and fitness-for-service evaluations will provide the best pathway forward to obtain the greatest yield from resources in terms of increasing pipeline safety and reliability and providing the largest return on investment. As outlined in **FIG. 2**, consistent and repeatable approaches (technical building blocks and programs) that can easily integrate and aggregate into solutions for more complex compliance and technical circumstances will serve to best support this pathway. On the left of the figure are some primary and specific conditions and threats that must be addressed. In the middle of the figure is where the technical building blocks are improved or developed. The far right of the figure shows the programs that will be constituted by a logical and pertinent arrangement of the technical building blocks to best achieve compliance and, most importantly, improve safety and reliability performance. These programs will be composed of consistent and repeatable workflows.

Supporting the technical building blocks and pathway toward effectively demonstrating safe and reliable pipeline operations is a well-structured engineering assessment approach and document design that provides reviewing stakeholders with a logical flow of technical analysis in support of the engineering assessment conclusions as outlined in the basic example in **FIG. 3**.

Code-related considerations. In the Canadian standard, the requirements for engineering assessments are provided in clauses 3.4 and 10.1 of CSA Z662. Clause 3.4 establishes the structural requirements for the engineering assessment, while Clause 10.1 provides the detailed elements that need consideration as applicable. It is noted that there is an opportunity to focus the engineering assessment as applicable to a specific threat or circumstance, but there is also a requirement to consider risk assessment as part of the analysis.

With respect to the U.S. CFR and Gas Mega Rule requirements, 49 CFR Part 192.632 (Engineering Critical Assessment for Maximum Allowable Operating Pressure Reconfirmation: Onshore Steel Transmission Pipelines) requires that, when operators conduct an MAOP reconfirmation in accordance with 49 CFR Part 192.624(c)(3) Method 3, using an engineering critical assessment to establish the material strength and MAOP of the pipeline segment, the engineering critical assessment must assess the following: *threats; loadings and operational circumstances relevant to those threats, including along the pipeline right of way; outcomes of the threat assessment; relevant mechanical and fracture properties; in-service degradation or failure processes; and initial and final defect size relevance. The engineering critical assessment must quantify the interacting effects of threats on any defect in the pipeline.*

Of particular interest, as it pertains

Table of Contents

1 INTRODUCTION AND BACKGROUND 3
2 OBJECTIVE AND SCOPE 3
3 PIPELINE OVERVIEW 3
3.1 Pipeline Configuration and Operating Parameters 3
3.2 Pipe Properties 5
4 PIPELINE HISTORY 7
4.1 Incident History 7
4.2 Operating History 7
4.3 Pipeline Integrity Management History 7
4.3.1 Pressure Testing 7
4.3.2 Integrity Assessments 7
4.3.3 Integrity Excavations 8
4.4 Class Location 8
5 THREAT ASSESSMENT 8
5.1 External Corrosion 8
5.2 Internal Corrosion 16
5.3 Cracking (Stress Corrosion Cracking) 17
5.4 Deformation 28
5.5 Geotechnical and Hydrotechnical 31
5.6 Operational 32
5.7 Materials and Construction 32
5.8 Interacting Threats 33
6 RISK ASSESSMENT 34
6.1 Risk Assessment Results 34
6.2 Evaluation of Risk Results and Risk Control 39
7 CONCLUSIONS 39

FIG. 3. Example of engineering assessment document structure.

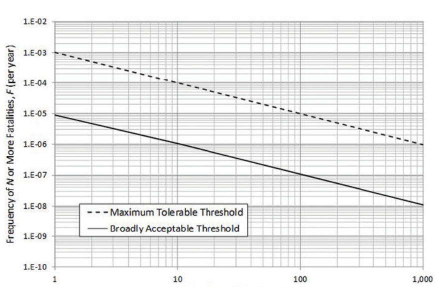


FIG. 4. CSA Z662:23 proposed societal risk acceptance criteria.

to a potential pathway forward for demonstrating margins of safety for existing pipelines vs. pipeline replacement, the CSA Z662 Risk Management Task Force has developed quantitative safety risk and environmental risk acceptance guidelines that are proposed for the 2023 standard (at the time of this publication, this standard has not yet been approved) as part of the non-mandatory Risk Assessment Annex (Annex B) in the CSA Z662 standard. **Note:** Non-mandatory reliability-based targets, as a function of safety and environmental consequence factors for the design of natural gas and low vapor pressure (LVP) liquid hydrocarbon pipelines, respectively, are provided in Annex O of the current CSA Z662:19 standard. These risk acceptance thresholds are expected to be a useful tool in demonstrating pipeline safety to stakeholders in the context of both pipeline risk assessments and engineering assessments. The proposed risk acceptance criteria are as follows:

- Guidance on safety risk (likelihood of failure × consequence of failure) acceptance criteria:
  - Individual risk (i.e., the annual probability of fatality for an individual situated at a particular location): The broadly acceptable individual risk threshold is 1×10^{-6}/yr, and the maximum tolerable risk threshold is 1×10^{-4}/yr.
  - Societal risk (i.e., a measure of the risk where the consequence considered is a function of the expected number of fatalities): The broadly acceptable and maximum tolerable thresholds are as established on an *F-N* curve, as shown in FIG. 4. Societal risk is typically presented in the *F-N* curve, which plots the cumulative frequency of events that can cause *N* + number of fatalities vs. number of fatalities.
- Guidance on environmental risk acceptance criteria:
  - The broadly acceptable environmental risk threshold is 2.4×10^{-2} m^3/km-yr, and the maximum tolerable risk threshold is 2.4 m^3/km-yr (the impact-adjusted release volume). The impact-adjusted release volume is estimated with due consideration of factors, including failure mode and associate pipe opening, pipeline diameter, flowrate, product properties, sectionalizing valve spacing, release detection, isolation time and environmental impact based on location.

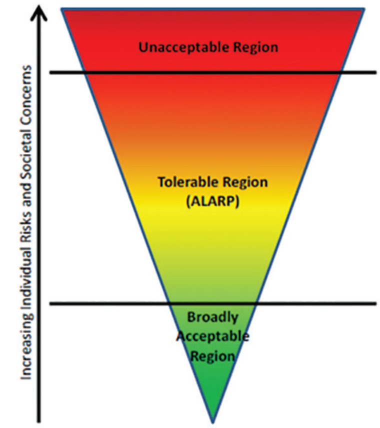


FIG. 5. The ALARP concept.

With respect to the broadly acceptable and maximum tolerable risk thresholds noted in the above criteria, the region between these two thresholds is considered tolerable risk if it can be demonstrated to be as low as reasonably practicable (ALARP), as shown in FIG. 5. The triangle represents the decreasing risk and the diminishing proportional benefit as risk is reduced.

Demonstration of ALARP requires that the cost of further risk reduction is grossly disproportionate to the risk reduction achieved. So, with respect to risk acceptance thresholds, if it can be demonstrated that the risk levels satisfy broadly acceptable or ALARP requirements, a compelling case for pipeline safety can potentially be made in the context of a pipeline engi-

neering assessment. The application of risk acceptance criteria, such as those proposed for the CSA Z662:23 standard, provides a strategic framework for improving pipeline safety and the use of risk assessment within an engineering assessment.

Quantitative risk acceptance criteria are also of interest to the U.S. regulatory bodies, and have also been explored through recent PHMSA industry studies and work groups related to quantitative risk modeling and risk tolerance thresholds. Independent risk and reliability-based safety cases are increasingly being requested as an important element of engineering assessments by U.S. regulatory bodies.

Data integration. Accurate data underlies the successful implementation of engineering assessments. This includes all pipe material attributes, construction, inspection and operating data necessary for the engineering assessment/fitness-for-service frameworks for all relevant threats and conditions.

Creating data integration roadmaps for each technical building block and aggregating these on a pipeline system basis yields a vision of the required data flows to support an effective engineering assessment. This vision includes being able to extrapolate learnings across sampling populations and pipeline systems, and across the industry in data warehousing approaches. Data integration requirements would be documented in standard operating procedures and fully integrated into an operator's business management infrastructure (such as a geographic information system, a warehouse management system and/or data management software). The current focus on and requirements for materials verification and traceable, verifiable and complete records align with these data needs. Minimizing the use of default values ensures that a risk-prioritized focus is not distorted.

Integrity data collected from continuously evolving and improving ILI technologies will increasingly act as a key foundation for engineering assessments, including ILI tools for metal loss, cracking, mechanical damage and specific conditions (i.e., hard spots, long seam flaws and pipe movement). Therefore, it is important that the objectives are clearly defined, and that the correct ILI technology is applied for assessing the applicable threats. This includes consideration of the detection and sizing capabilities. For example, the newer-generation electromagnetic acoustic transducer (EMAT) ILI technology has evolved significantly to a point where it can play a primary role in stress corrosion cracking assessment and management programs for gas transmission pipelines. The National Association of Corrosion Engineers (NACE) International SP0102 standard,[4] as referenced in the American Petroleum Institute (API) 1163 standard,[5] provides guidance on proper tool selection by threat, as shown by the excerpt in FIG. 6.

Types of ILI Tools and Inspection Purposes[(7)]							
Anomaly	Imperfection/Defect/Feature	Metal Loss Tools		Crack Detection Tools			Deformation Tools
		Axial MFL	Ultrasonic Compression Wave[(A)]	Ultrasonic: *Liquid Coupled*[(A)]	Ultrasonic: *EMAT*	Transverse MFL	
Metal Loss							
	External Corrosion	Detection, Sizing	Detection, Sizing	No Detection	No Detection	Detection, Sizing	No Detection
	Internal Corrosion						Limited Detection[(D)], Sizing[(D)]
	Gouging	Detection[(B)], Sizing[(B)]	Detection[(C)], Sizing[(C)]			Detection[(B)], Sizing[(B)]	No Detection
Crack Like Anomalies							
	Narrow Axial External Corrosion	Detection	Detection, Sizing	Detection, Sizing	Detection, Sizing	Detection, Sizing	No Detection
	Stress Corrosion Cracking (SCC)	No Detection	No Detection	Detection, Sizing	Detection, Sizing	Limited Detection,[(E)] Sizing	No Detection
	Fatigue Cracks	No Detection	No Detection	Detection Sizing	Detection, Sizing	Limited Detection,[(E)] Sizing	No Detection
	Long Seam Cracks, etc. *(toe cracks, hook cracks, incomplete fusion, preferential seam corrosion)*	No Detection	No Detection	Detection, Sizing	Detection, Sizing	Detection[(E)] Sizing	No Detection
	Circumferential Cracks	Limited Detection,[(E)] Sizing	No Detection	Detection[(F)], Sizing[(F)]	No Detection	No Detection	No Detection
	Hydrogen-Induced Cracking (HIC)	No Detection	Detection, Sizing	No Detection	No Detection	No Detection	No Detection
Weld Anomalies							
	Lack of fusion in LW	No Detection	No Detection	Detection, Sizing	Detection Sizing	Detection[(E)], Sizing[(E)]	No Detection
	Lack of fusion in GW	Detection[(E)] Sizing	No Detection	Detection[(F)], Sizing[(F)]	No Detection	No Detection	No Detection
	Girth Weld Anomaly *(voids, etc.)*	Detection, Sizing	Detection, Sizing	Detection[(F)], Sizing[(F)]	No Detection	No Detection	No Detection
Deformation							
	Sharp Dents	Detection[(G)]	Detection[(G)]	No Detection	No Detection	Detection[(G)]	Detection, Sizing
	Flat Dents	Detection[(G)]	Detection[(G)]	No Detection	No Detection	Detection[(G)]	Detection, Sizing
	Buckles	Detection[(G)]	Detection[(G)]	No Detection	No Detection	Limited Detection[(G)]	Detection, Sizing

FIG. 6. Excerpt from ILI System Selection Guidance in NACE SP0102.

A documented verification and validation assessment of the ILI tool performance is a necessary foundation for using the ILI data as part of the engineering assessments. Also, qualification programs related to non-destructive evaluation tools used in the field for such validation also play a key role in ensuring that the ILI results are accurately validated.

Where ILI data serves as a foundation for the engineering assessment, it should not be taken for granted that all stakeholders are familiar with, or accepting of, the ILI technology applied (particularly for less-established technologies). A full API 1163-based analysis, including the tool selection against the objective, the run verification (i.e., data quality assessment, run functionality checks) and the tool performance validation (i.e., unity plots of tool vs. infield measurements) should be included. There should also be workflows and analysis steps to deal with outliers in tool performance. There must be pathways for investigation and root cause analysis for outliers, and, where necessary, opportunities to reject the inspection where outliers or significant data degradation gaps cannot be resolved.

Part 2. Part 2 will be published on www.GasProcessingNews.com. **GP**

NOTES

This article was first presented at the 34th Pipeline Pigging & Integrity Management Conference, February 2022.

[a] On October 1, 2019, the PHMSA finalized a rule for gas transmission pipelines that had been in development for more than 8 yr. The effective date is designated as July 1, 2020. This highly anticipated rule (referred to as the Gas Mega Rule) is the most significant change to existing gas pipeline regulations since 1970 and is likely to have a significant impact on the operation of gas transmission assets in the U.S. The rule is split into three parts, generally focusing on MAOP reconfirmation, material verification and integrity assessments outside of HCAs (Part 1); improvements to repair and response criteria and to the integrated master plan (IMP) (Part 2, still anticipated); and gas gathering (Part 3, released in November 2021).

LITERATURE CITED

Complete literature cited available online at www.GasProcessingNews.com.

BEN MITTELSTADT is Director of Technical Services at Dynamic Risk Assessment Systems Inc.

DAN WILLIAMS is a Principal Consultant at Dynamic Risk Assessment Systems Inc.

How digital technology helps processing industries connect the dots

R. FARR, GE Digital, Houston, Texas

Leaders in the processing industry are exploring ways of using digital technology to create intelligent and autonomous operations that optimize daily production, manage risk, improve margins and enable a more flexible workforce.

Gas turbines play a critical role across the processing industries, such as mechanical drive applications in LNG facilities, compressor stations on natural gas pipelines and cogeneration power utility islands within a processing facility. Digitizing gas turbines and the associated balance of plant equipment can provide significant payback in response to continued pressure to drive down costs, increase efficiency, and improve reliability and availability while meeting new regulations to reduce emissions.

Connecting the dots. By combining closed- and open-loop optimization, whether the producer's objective is to meet profitability key performance indicators (KPIs) or sustainability goals, digitizing gas turbines makes it possible to connect the dots and meet those goals. As such, producers can extract more value from existing assets with software by bringing intelligent and autonomous operations to bear in three ways:

1. Deploy new operating modes in gas turbines and plant controls that automate and enable flexible, reliable operations
2. Optimize dispatch by ensuring that operations and commercial teams have the best available information on asset- and plant-specific capabilities and costs to manage risk, uncertainty and improve margins
3. Maintain optimal performance across changing operations by monitoring thermal performance and flexible operating modes using plant-level diagnostics to quickly identify and cost-effectively address performance shortfalls.

Digital technologies can help processors connect the dots to make their gas turbines and facilities more adaptable to meet production goals and reduce risk and cost. Software can provide real-time insights, powerful optimization and data-informed decisions. By deploying new operating modes, optimizing dispatch and maintaining optimal performance, digitization helps producers decrease fuel costs, meet emissions regulations and improve availability to drive better economic outcomes.

Deploy new operating modes. Producers in the processing industries increasingly need tools that help reduce the cost of operations, but not at the expense of reliability and availability. New digital technologies have shown the ability to be deployed for gas turbine controls that enable flexible operations and process automation while improving the costs of operation and reliability. An example of such technology is applying artificial intelligence (AI) and machine-learning (ML) to the tuning of gas turbine combustion.

Gas turbines typically require seasonal adjustment, a tuning or mapping of flame temperatures and fuel splits to enable dependable and emissions-compliant operations as weather patterns change throughout the year. This is a manual process performed by an onsite expert and requires an outage that impacts availability, resulting in operational challenges in the processing industry where gas turbine uptime is critical to meeting overall production goals. Furthermore, manual seasonal tuning is only efficient for the precise conditions under which it was completed and does not enable the gas turbine to respond effectively to ambient temperature or fuel property changes between tunings.

Demonstrations have shown that by digitizing the gas turbine and utilizing AI/ML technology to continuously optimize combustion in closed-loop control—in place of manual seasonal adjustments—aero-derivative gas turbine operators can achieve the following benefits:

- 0.5%—1% reduction in fuel consumption/carbon dioxide (CO_2) emissions
- Up to 14% reduction in carbon monoxide (CO) emissions
- Up to 12% reduction in nitrogen oxide (NO_x) emissions
- Improved availability with no manual tuning or associated downtime.

By deploying AI-enabled tuning software in a supervisory control system fully bounded by control system safety-critical programming, ML can be used to safely determine the ideal flame temperatures and fuel splits continually and autonomously for optimal combustion, considering critical variables such as ambient conditions and fuel quality change. By sensing changes in ambient temperature, gas fuel properties and degradation, real-time adjustments can be sent to the turbine controls in seconds to deliver cleaner, more efficient and reliable operations. Another benefit of letting software do the tuning is that it allows experts to be deployed to address more critical actions.

Optimizing dispatch. Once a gas turbine has been digitized and enabled with new operating modes, digital technology can provide additional value by optimizing the dispatch of the plant and gas turbines. Many gas turbines are used for cogeneration in a power utility island of a processing facility. Meeting flexible and some-

times conflicting thermal and electrical demands is needed in such applications.

Plant operators have traditionally lacked comprehensive, timely and accurate data and tools that provide a single integrated source of truth to make sense of it all. Only with a complete picture of a plant's operations, maintenance and actual capability, coupled with market and environmental conditions, is it possible to make decisions about dispatch that maximize revenues while reducing fuel and maintenance costs.

This is where digitization comes in. Software can provide the information and analytics necessary to operate and dispatch a plant and gas turbine in a way that meets today's complex and continuously evolving market dynamics most efficiently. For example, AI-enabled software can reduce dispatch plan uncertainty with accurate capacity predictions (power and steam) and heat rate for the gas turbines and other supplementary power assets in the plant, such as duct burners, boilers and inlet conditioning. When coupled with a unit commitment optimizer using advanced numerical methods that can manage the wide range of constraints and inputs required, fuel and CO_2 savings of 0.5% or more may be achievable.

As weekly or daily plans change during operations due to a shift in demand or ambient condition, a digitized plant can respond almost instantly, without any manual handoffs, to optimize based on the latest inputs and constraints.

Maintain optimal performance. In addition to deploying new operating modes and optimizing dispatch, digital technology can be applied to maintain the optimal performance of the gas turbine and plant. Assets (i.e., gas turbines) degrade over time, and companies are always searching for ways to keep their assets healthy. To do that, companies must have visibility into the operation of that asset, not only to prevent unplanned and unwanted downtime, but to recover lost performance. This recovered degradation will reduce the cost of fuel, save emissions and potentially boost output. Operational and system efficiencies are critical to ensure business and environmental performance are paramount.

Industry professionals responsible for managing assets are faced with numerous challenges in designing cost-effective maintenance strategies for equipment. Intelligent approaches for identifying optimal processes and systems for managing assets combine data-driven modeling of known degradation mechanisms with field expertise in asset performance and plant operations.

Many companies use digital twins, digital representations of physical assets that apply advanced analytics and ML to reduce operational costs and risks. Digital twins are a key piece of the digital transformation puzzle and a great example of how software connects the dots. By embedding digital twins in fully automated software, engineering and plant personnel can achieve operating targets through benchmarking, analytics-based alerts and advisors across thermal performance and flexible operating modes, such as starts, ramping and minimum load.

Takeaway. Without digitizing a gas turbine or plant, ensuring the assets are operating as expected can be challenging. For example, a gas plant's primary operating cost is fuel. Heat rate—a measurement of plant efficiency—is a moving target that changes with seasonal load profile, operating modes, ambient conditions and equipment health. Heat rate is not directly measurable in the plant, so without a digital twin to benchmark performance against, it can be unclear whether a change in heat rate is attributable to a change in operating conditions or a change in equipment health and degradation. With clarity of the source and the magnitude of the change, the most efficient use of resources can be deployed to achieve the most significant impact on recoverable degradation and ensure optimal performance over time.

Automating more operations creates a more reliable and safe plant. To improve profitability and reliability while lowering operation and maintenance costs, plant operators must evolve and adapt. Tapping digital tools that enhance flexibility, visibility and analytics for optimal operation is one readily achievable way to do that. **GP**

RACHEL FARR is a Senior Director of Product Management for GE Digital, and is responsible for delivering operations performance management software solutions that increase revenue and margins with visibility, insights, decision support and advanced edge controls to optimize performance for power generation fleets. Ms. Farr has more than 18 yr of experience in digital power and aviation, gas turbine and plant technologies, and product and technology strategies. Ms. Farr holds a BS degree in materials science and engineering from the University of Pennsylvania and an MS degree from Ohio State University.

Considerations for not providing a dedicated spare pressure-relief device

P. K. NAMILAKONDA, Samsung Engineering, Houston, Texas

Pressure-relief devices (PRDs) play a critical role in the hydrocarbon processing industry. They act as the last line of defense in protecting plant equipment from overpressure and are mandated by the American Society of Mechanical Engineers (ASME). This article provides some key factors that process engineers and plant operators must consider to determine if an installed spare PRD is necessary.

Like many other plant components, relief devices require routine maintenance to ensure safe and reliable plant operation. While most of the PRDs in hydrocarbon processing plants can be taken out for maintenance during plant turnarounds, certain PRDs may require more frequent maintenance while the plant is in operation (onstream). The term "onstream" refers to the overall plant operation. Reasons for more frequent maintenance could include safety requirements or possibly because the PRD is installed in a fluid service where fouling, corrosion or plugging could potentially impact PRD operation.

Onstream maintenance of a specific PRD in the plant can be achieved by one of the following options:

1. Providing an alternate and adequate relief protection to the protected equipment (e.g., a spare PRD, or an alternate and adequate flow path to another PRD)
2. Isolating, draining and depressurizing the protected equipment from all possible overpressure sources (taking the equipment out of service)
3. Eliminating as many overpressure sources as feasible, and implementing administrative controls for monitoring the protected equipment for the duration of the PRD maintenance.

Process engineers are required to review these options during the front-end design development step, with input from the plant operations team, to make an appropriate decision. PRDs requiring onstream maintenance are, in general, provided with isolation valves on inlet and outlet (if applicable) pipelines. ASME code requires administrative controls to be in place for proper operation of the isolation valves in the relief path so the equipment's overpressure protection is not compromised. These requirements include providing locks or car seals in addition to administrative controls to prevent accidental closure (for more information, refer to ASME BPVC Section XIII Non-Mandatory Appendix B). Process engineers and plant operators must be familiar with the ASME code requirements, and Appendix B provides valuable guidance and requirements for determining a need for installing a dedicated spare PRD. The following will review the three options in further detail.

Option 1. Providing an installed spare PRD is generally the preferred way to provide adequate overpressure protection to the process equipment. While this option involves initial capital expenditure for the PRD and associated piping, it also provides the safest way to keep the protected equipment in continuous operation for the duration of the maintenance. Plant operators are required to follow stringent administrative procedures for isolating the PRD requiring maintenance.

In the absence of an installed spare PRD, other creative ways are available to achieve the same purpose. This includes utilizing PRDs on nearby equipment to provide the necessary protection by means of utilizing an existing pipe routing for a temporary relief path. However, process engineers should perform a thorough analysis to ensure that such a temporary and alternate relief path is adequate in terms of hydraulic requirements (inlet pipe pressure loss), PRD flow area and fluid service compatibility.

Option 2. This option is best suited for equipment that can be taken out of service while onstream plant production is maintained. A spare PRD may not be required if the entire protected system can be isolated during normal plant operation. Equipment operating in intermittent service also may not require a spare PRD. For example, consider a hydrocarbon system containing multiple catalyst bed vessels in parallel, where at least one vessel is always isolated for catalyst regeneration and is in standby mode for a prolonged duration. A spare PRD may not be provided for protecting such vessels, as maintenance can be performed when the vessel is in standby mode. In such cases, the duration of the maintenance should be given careful consideration, as the vessel cannot be returned to service without code-compliant relief protection. Plant operators generally rely on administrative procedures to open and close several valves for placing equipment back in service. Failure to properly follow these administrative procedures can compromise the overpressure protection mandated by the ASME code. One such safety incident was the ruptured heat exchanger in June 2008 that killed one worker and injured six others at the Goodyear Tire and Rubber Co. in Houston, Texas.[1] One of the root causes of this accident was failure to properly follow the procedure of opening the PRD inlet piping isolation valve before placing the heat exchanger into service.

PRDs requiring onstream maintenance are, in general, provided with isolation valves on inlet and outlet (if applicable) pipelines. ASME code requires administrative controls to be in place for proper operation of the isolation valves in the relief path so the equipment's overpressure protection is not compromised.

Option 3. This is the least preferred option, as it involves operating the process equipment without overpressure protection under administrative controls. One of the first things that the plant operator is required to do in this option is to identify all potential overpressure sources beforehand and attempt to eliminate or minimize the likelihood of overpressure from such sources for the duration of the PRD maintenance. In this option, assuming that isolation valves are in place, the PRD must first be isolated by closing the inlet and outlet isolation valves before the maintenance can begin. Closure of the isolation valves would lead to blocking the relief path and to operating the protected equipment with a risk of overpressure. Plant operators are required to stay in compliance with the ASME code for proper implementation of administrative controls. ASME BPVC-XIII Appendix B-7(d) states:

> "Procedures are in place to provide pressure relief protection during the time when the system is isolated from its pressure relief path. These procedures should ensure that, when the system is isolated from its pressure relief path, an authorized person should continuously monitor the pressure conditions of the vessel and should be capable of responding promptly with documented, pre-defined actions, either stopping the source of overpressure or opening alternative means of pressure relief. This authorized person should be dedicated to this task and should have no other duties when performing this task."

Plant operators should consider several factors to safely rely on administrative controls for monitoring unprotected equipment. One such factor is the duration of maintenance. Can the relief device be returned to service in a reasonable amount of time? Another factor is training. What kind of training would be required for authorized persons to respond promptly in an event of pressure excursion in the process service? If possible, process engineers and plant operators should steer away from this option, as this involves several factors that cannot be fully accounted for during the plant design.

Multiple overpressure scenarios. There can be more than one overpressure scenario that is applicable for a PRD. Overpressure due to an external pool fire is a well-known scenario that is applicable for process equipment at grade level. Additional overpressure scenarios include blocked outlets and auto-control failures, among others (see API 521 for more information). The process engineer should review all applicable overpressure scenarios when determining the need for a spare PRD.

It is a generally accepted industry practice not to provide spare PRDs in the following applications, as the PRD maintenance interval typically aligns with the protected system maintenance interval:

- PRDs installed exclusively for protecting equipment from external fire (fire-only PRD)
- PRDs installed in cooling water service exclusively for protecting the equipment from thermal expansion (cooling water thermal).

Multiple PRDs. Multiple PRDs are often provided when a single PRD is inadequate to handle the required relief flowrate. The ASME code allows staggering the PRD set pressures in this case. For multiple PRDs:

- A single spare PRD may be installed with the basis that administrative controls are in place to perform maintenance on one PRD at a time. When a single spare PRD is installed, performing maintenance on multiple PRDs would result in inadequate relief protection, which is a possible code violation.
- The spare PRD should have an orifice flow area identical to the operating PRD with the largest area.
- When the spare PRD is put into operation, at least one PRD should have a set pressure at 100% of the maximum allowable working pressure.

Takeaway. The ASME code does not mandate installing a spare PRD. However, in the absence of an alternate and adequate relief path, the ASME code mandates the implementation of proper administrative controls for safe isolation and subsequent monitoring of the protected system for the duration of PRD maintenance. Administrative controls play a critical role in the safe operation of any plant, and many plant operators prefer to install a spare PRD to provide an adequate and equivalent overpressure protection. Process engineers and plant operators must be conversant with the applicable regulatory requirements and must review all PRDs on a case-by-case basis to determine whether an installed spare PRD is, or is not, necessary. HP

LITERATURE CITED

1. U.S. Chemical Safety and Hazard Investigation Board, "Goodyear heat exchanger rupture," January 2011 (report released), online: https://www.csb.gov/goodyear-heat-exchanger-rupture/
2. ASME Boiler and Pressure Vessel Code, Section XIII-2021, "Rules for overpressure protection."

PRANAY K. NAMILAKONDA is a Process Engineer with Samsung Engineering America, a global EPC firm based in Houston, Texas. He is a licensed Professional Engineer and has more than 15 yr of experience in front-end engineering execution of various hydrocarbon projects. He also leads the process engineering group, with an emphasis on engineering automation. Mr. Namilakonda earned an MS degree in chemical engineering from Oklahoma State University, and a BTech degree in chemical engineering from Osmania University, India.

Decarbonize LNG liquefaction with pre-combustion CO_2 capture technology

D. VESKOVIC, J. BEARD, M. ROBERTS and D. GRAHAM,
Air Products and Chemicals Inc., Allentown, Pennsylvania

In recent years, lifecycle greenhouse gas (GHG) emissions studies of the natural gas industry have drawn attention to the significant environmental impact of liquefied natural gas (LNG). Not only is carbon dioxide (CO_2) released when natural gas is burned at the end user, but the processes of extracting, treating, liquefying, shipping and regasifying the natural gas all contribute to the total emissions of using natural gas as fuel. With natural gas representing a rising fraction of the world's total energy consumption, the importance of reducing these emissions will continue to grow in the future.

Emissions from end users occur across many different geographic locations, from large-scale power generation facilities to individual homes. For many of these users, capturing and sequestering the CO_2 associated with their emissions is challenging for various reasons (economics, lack of sequestration infrastructure, etc.). Conversely, LNG liquefaction facilities present a unique opportunity to capture GHG emissions, as they provide a single location where a significant portion of emissions is concentrated.

For both retrofits to existing plants and for new-build plants, pre-combustion CO_2 capture can significantly reduce the overall carbon footprint of LNG. This article examines how a proprietary approach to implementing and integrating pre-combustion capture with an LNG liquefaction facility can reduce the overall CO_2 emissions from a liquefaction facility by more than 97%. This approach comprises the following three main steps:

1. Producing hydrogen (H_2) fuel from natural gas or other fuel gases within the liquefaction facility
2. Removing CO_2 from the syngas generated during H_2 production
3. Using the H_2 as the primary fuel in the plant's gas turbines.

The details of this configuration are flexible and can be adjusted to serve individual project needs.[1–5]

Methods to reduce CO_2 emissions. CO_2 emissions at an LNG liquefaction facility can be categorized into three primary sources:

1. CO_2 vented during upstream pretreatment to remove acid gases
2. CO_2 released in the flue gas from gas turbines used to power the liquefaction process
3. CO_2 released in the generation of power for the remainder of the facility.

Two broad approaches can be utilized to reduce the CO_2 emissions from these sources. One approach is to reduce the amount of CO_2 generated, primarily by improving efficiency. The second is to capture and sequester the CO_2 in the feed and the CO_2 created during power generation. A combination of both approaches is necessary to achieve a high degree of overall emissions reduction.

Capturing and sequestering the CO_2 in the feed is already possible with well-proven technologies. A typical acid gas removal unit (AGRU) for LNG already separates CO_2 from the natural gas feed using an amine-based absorption process. Although many facilities today reject that CO_2 to atmosphere, some already possess the capability to capture CO_2 removed this way and sequester it. The use of this approach can reduce the total CO_2 emissions from an LNG facility by up to 30%, depending on the amount of CO_2 found in the feed.

Improvements to plant efficiency can also reduce CO_2 emissions. Although many LNG facilities use simple-cycle industrial gas turbines, newer facilities have made use of aeroderivative gas turbines or combined-cycle arrangements to improve power efficiency. With improved power efficiency, less fuel is required to generate the same power available for liquefaction. This results in lower emissions for the same LNG production. The use of aeroderivative turbines can potentially reduce the CO_2 emissions of a facility by approximately 14% over equivalent industrial gas turbines, while a combined-cycle arrangement can reduce CO_2 emissions by approximately 25% over a simple-cycle arrangement.

While these existing methods can significantly reduce the amount of CO_2 released from a facility, they cannot address all the CO_2 released in the flue gas from a plant's gas turbines. To capture the remaining CO_2, a different approach is required.

Flue gas CO_2 reduction via pre- or post-combustion capture. Two potential approaches to reduce CO_2 emissions from methane-burning gas turbines are post-combustion capture and pre-combustion capture.

With post-combustion capture, shown in **FIG. 1,** air and methane-rich fuel are fed to the gas turbine and CO_2 is directly removed from the flue gas by a CO_2 capture system. In this configuration, CO_2 is available at low partial pressure (near-atmospheric pressure with single digit mol% of CO_2) and high temperature. These conditions make it challenging to design a cost-effective recovery system that captures most of the CO_2. Furthermore, in retrofit applications, the lack of plot space close to the gas turbines can be problematic. The low pressure and high volume of flue gas may be impractical to transport to a distant CO_2 recovery unit.

Conversely, pre-combustion capture, shown in **FIG. 2,** captures CO_2 from streams at higher pressure and with higher fractions

of CO_2 (20%–30%). Pre-combustion capture needs a far smaller CO_2 recovery unit but requires other additional equipment to process the hydrocarbon fuel gas into clean H_2 for use in the gas turbines. Although not directly used for LNG service, the required technology for pre-combustion capture has been used for decades. For example, the authors' company's Port Arthur CO_2 capture project has captured approximately 1 MM metric tpy of CO_2 from two large-scale H_2 plants since it commenced operations in early 2013.

The advantages of pre-combustion capture allow for easier retrofit into existing plants and easier siting of the additional equipment needed for CO_2 capture. The proprietary pre-combustion capture process[a] uses a pre-combustion approach to capture the CO_2 that would be released in the gas turbine flue gas.

Pre-combustion capture technology overview. The proprietary process consists of three major systems (shown in **FIG. 3**):

1. The LNG liquefaction plant, including pretreatment
2. The H_2 production facility
3. The power generation system.

The LNG liquefaction plant takes in raw natural gas feed. Gas pretreatment, including acid gas removal, mercury removal and any heavy component removal, occurs upstream of the liquefaction plant. The liquefaction plant produces LNG from the pretreated gas while also producing a fuel stream that consists of end-flash and boil-off gas. Normally, this fuel stream would be sent directly to the gas turbines to generate power. However, in the proprietary pre-combustion capture process, the fuel gas stream, which is primarily methane (CH_4) and nitrogen (N_2), is instead sent to an H_2 production plant. The specific technology used for H_2 production can vary; potential options are discussed below.

In the H_2 production facility, the fuel gas stream is combined with oxygen (or air) and steam. The combined stream is sent to a series of reactors to generate a syngas stream of primarily H_2 and CO_2. The syngas stream contains CO_2 at high pressure (> 30 bar) and at a high fraction of the total stream molar flow (20%–30%). These conditions are well suited for CO_2 separation from H_2 product at moderate cost.

CO_2 is removed at two points in the process. First, the CO_2 in the natural gas feed is removed in the AGRU. Next, CO_2 generated during the reforming and water-gas shift reactions within the H_2 production facility is removed during a downstream separation step. This separation can be achieved through an amine unit similar to the AGRU, or through vacuum swing adsorption (VSA), cryogenic distillation or other techniques. Several of these technologies can achieve the target of > 97% CO_2 recovery.

After CO_2 removal, H_2 from the production facility is combined with CH_4 or used as pure H_2 fuel. The fuel is mixed with one or more diluents to suppress nitrogen oxide (NO_x) formation and is sent to the power generation system, which produces power for liquefaction and the H_2 production facility.

The power generation system may be simple or combined-cycle, and it may either be directly coupled to the liquefaction compressors or instead coupled to generators to produce electricity used by electric motors in the liquefaction facility.

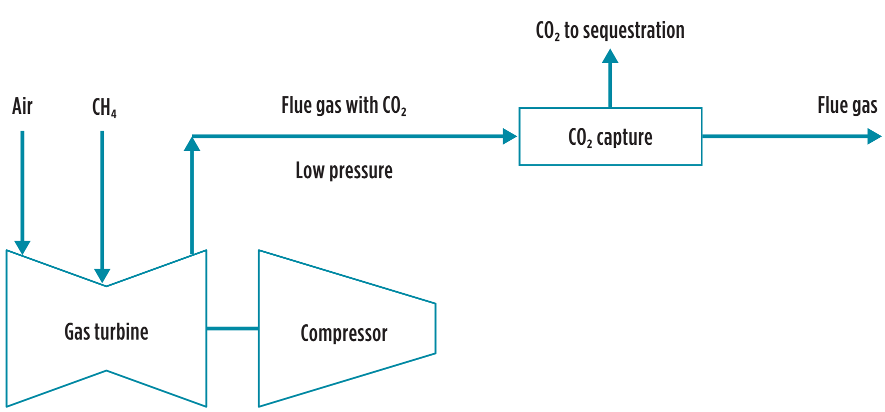


FIG. 1. Generic configuration for post-combustion CO_2 capture.

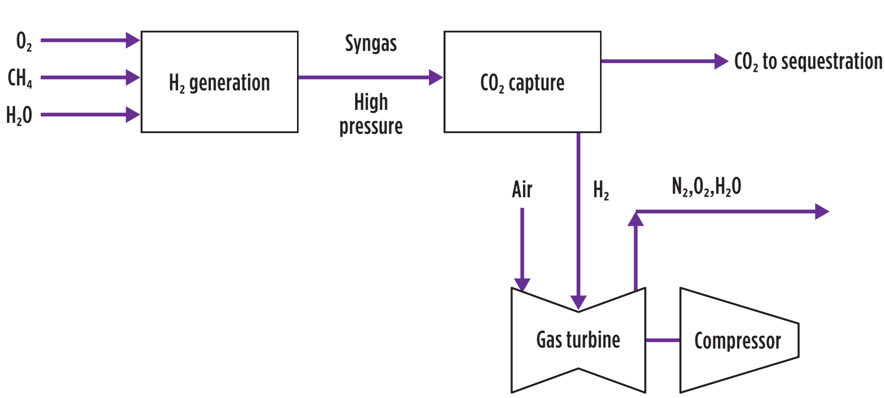


FIG. 2. Generic configuration for pre-combustion CO_2 capture.

Hydrogen production facility options. The pre-combustion capture process[a] can use a wide variety of technologies for its H_2 production plant, including the steam methane reforming (SMR) process, the autothermal reforming (ATR) process, or the partial oxidation (POX) process. All three options include a main reactor, one or more water-gas shift reactors, and several heat exchangers for steam generation.

The SMR process uses a tubular catalytic main reactor where fuel is burned at low pressure externally to the tubes to provide the heat needed for the reforming reaction. The SMR process is an industry standard for gray H_2 production thanks to its energy efficiency, high reliability and attractive capital cost. However, because fuel is burned at low pressure, approximately 40% of the total CO_2 generated in the H_2 production facility is released in low-pressure flue gas and cannot be recovered through pre-combustion capture.

The POX and ATR processes instead provide the required heat for their reactions by reacting a portion of the feed stream within the reactor vessel. This allows all CO_2 formed during the reactions to remain within the high-pressure

stream exiting the reactor. The CO_2 can then be removed from the product stream by an amine system or other separation technology, enabling a high degree of carbon capture for both processes. The differences between the two technologies lie with the specific reactions involved and the type of reactor needed. The POX process uses a non-catalytic main reactor and uses steam for temperature control, while the ATR process uses a catalytic main reactor with steam consumed as part of the reaction.

TECHNOLOGY SYNERGIES

The pre-combustion capture process[a] features several synergies gained by integrating liquefaction, H_2 production and power generation. These synergies would be unavailable if a plant were designed to run on green electricity from a local grid, or if it imported H_2 for its gas turbines.

Improved liquefaction efficiency. The proprietary process allows for a re-optimization of the LNG liquefaction plant to improve efficiency. Higher fuel requirements for the H_2 production facility allow for increased end-flash flow from the liquefaction facility, potentially increasing LNG production by up to 2%. This may aid in debottlenecking existing facilities that would otherwise be limited by available liquefaction power.

Low gas turbine NO_x emissions. With H_2-burning gas turbines, a major concern is the generation of NO_x. The pre-combustion capture process has several features that help in maintaining equivalent levels of NO_x to a typical dry low emissions (DLE) system. In the process, the end-flash gas used to feed the H_2 production plant typically contains a significant fraction of N_2. This N_2 passes through the plant and is sent to the power generation system, where it serves as a diluent for the fuel and aids in suppressing NO_x formation. Excess steam and N_2 from the H_2 production facility are also available for use as additional diluents.

Retrofit considerations. The process can be applied at both new-build sites and in the retrofit of existing plants. Retrofitting an existing plant to decarbonize it comes with several unique challenges and opportunities, as discussed here.

Gas turbine capabilities. Many gas turbines used as compressor drivers for natural gas liquefaction can handle a fuel that consists of a mix of H_2 and CH_4, requiring only minor engine and package upgrades to ensure safety and reliability while appropriately handling the new fuel. However, for cases where the goal is a high degree of pre-combustion carbon capture, the gas turbine must be capable of handling fuel with a combustible component comprised of 100% H_2. Some turbines have already been designed for 100% H_2 fuel. For other existing turbines, running on this fuel requires a modification of the combustion system to manage the unique properties of high-H_2 fuels. In either case, the gas turbine and other site specifics—including emissions limits and utilities availability—must be considered when evaluating H_2 as a fuel.

Offsite H_2 facilities. At some sites, there is insufficient nearby plot space for an H_2 production facility. In these cases, it may be possible to export end-flash gas via a pipeline to a remote H_2 production facility and receive either H_2 fuel or electricity in return. Alternatively, an existing LNG facility may simply wish to replace a portion of its fuel with a decarbonized fuel from a third party without any major changes to its own facility. Purchasing decarbonized H_2 from an external source is a simple change that can decrease an LNG facility's carbon footprint.

CASE STUDIES

The following section examines three case studies implementing the proprietary pre-combustion capture process[a], two retrofit options for existing plants and a new-build that takes full advantage of both efficiency improvements and CO_2 capture options.

All three cases and the base case used for comparison assume an LNG liquefaction facility that uses a proprietary LNG technology[b]. This plant has net-in-tank LNG production of 4.5 MMtpy. The raw natural gas feed to the liquefaction plant pretreatment contains 4% CO_2, which is removed in the AGRU. Unless otherwise stated, the liquefaction power is provided by two directly coupled industrial gas turbines (Frame 7E) in simple-cycle, and an industrial gas turbine as a simple-cycle power generator for the electrical power requirements. For the purposes of this case study, the power generator can be another single Frame 7E, or multiple smaller gas turbines of similar efficiency and total power capacity.

For each case, the power used by liquefaction is held constant, while the total power was varied to address the power needs of the additional H_2 production facility.

Retrofit (partial and full) cases. Two options were evaluated for the retrofit case, differing on the degree of upgrades needed for the combustion system of the gas turbines. The "partial H_2" case sends a portion of the flash gas to the H_2 production facility, then recombines the H_2 product with the remainder of the flash gas. This resultant fuel stream contains approximately 20 vol% H_2. Existing industrial gas turbines designed for use with natural gas require only minor adjustments to handle this new fuel stream, which avoids a significant overhaul or extended downtime.

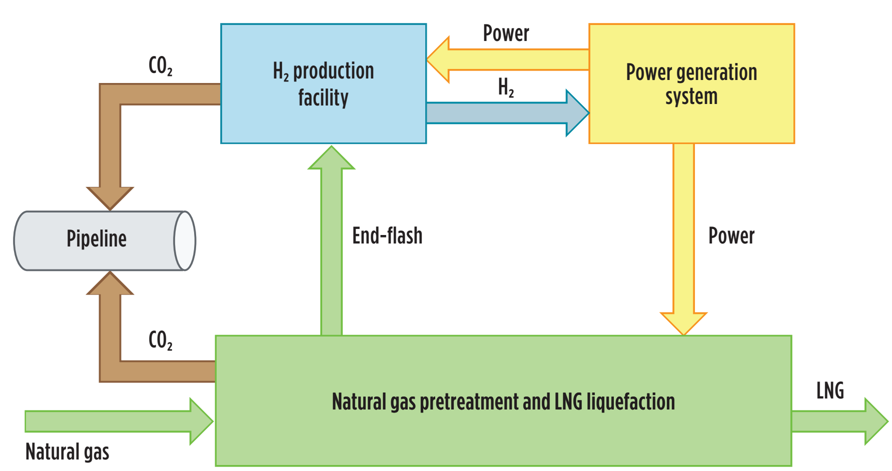


FIG. 3. Proprietary pre-combustion capture technology[a] layout overview.

TABLE 1. Comparison of retrofit and new-build CO_2 emissions to generic baseload LNG liquefaction plant

	Retrofit: Partial H_2	Retrofit: Full H_2	New-build
Gas turbine arrangement	Industrial GT, simple-cycle	Industrial GT, simple-cycle	Aero GT, combined-cycle
Natural gas feed vs. base case, %	100.7%	107.5%	99.1%
Net-in-tank production vs base case, %	100.4%	102.4%	99.5%
H_2 produced, MMft³d	18	299	176
Liquefaction specific power vs base case, %	99.6%	98%	100.5%
Auto-consumption vs. base case, %	103.3%	150.7%	96.8%
Total CO_2 captured, tph	75.4	297.6	197.8
% emissions reduction vs. base case	32.6%	96.6%	98%

The "full H_2" case sends all the flash gas to the H_2 production facility, generating a fuel stream of predominantly H_2 (with N_2 or steam as diluents). Although some industrial gas turbines are capable of handling such a fuel stream, these turbines may require a conversion of the combustion system and fuel delivery system modifications to do so.

In both options, additional gas turbine generators are required to supplement the existing electric power. These can be located at the H_2 production facility.

New-build case. The new-build case examines a potential new LNG liquefaction facility. This hypothetical plant utilizes three aeroderivative gas turbines in combined-cycle for power generation, as well as one or more small gas turbine generators to supplement the available power. Electric motors driving the liquefaction compressors use most of the generated power. The remaining power is used for the balance of plant. Liquefaction compressor power has been fixed at the value used for the base case for ease of comparison.

For all three cases, the following additional technology selections were made:

- High-conversion POX process for H_2 production
- Air separation unit (ASU) to provide low-purity gaseous oxygen for POX
- AGRU designed to recover CO_2 and compress to pipeline pressure
- Amine unit integrated into the H_2 production facility to separate CO_2 from the H_2 product.

The results for each case are shown in TABLE 1.

Takeaways. In the retrofit cases, the higher flash gas demand improved the liquefaction specific power—the liquefaction compression power required to produce 1 metric t of LNG—allowing more LNG to be produced for the liquefaction power. The partial H_2 case achieved 0.4% higher production, whereas the full H_2 case achieved 2.4% higher production. For the new-build case, a more efficient gas turbine selection reduced the flash gas requirement and slightly reduced the LNG production.

However, the new-build case's efficiency provides significantly better auto-consumption—the percentage of feed consumed as fuel—than the retrofit cases. The new-build case also requires a significantly smaller H_2 production facility for a similar liquefaction plant to achieve 97% emissions reduction. The overall CO_2 generated during H_2 production is also significantly lower for the new-build case compared to the full H_2 retrofit, which reduces the required size of the CO_2 compression system.

Both retrofit and new-build cases are capable of a high degree of overall CO_2 capture by capturing CO_2 from the natural gas feed and by pre-combustion capture of the CO_2 generated in the H_2 production facility.

The partial H_2 retrofit case achieved a CO_2 emissions reduction of 32.6%, an improvement over capturing CO_2 from the AGRU alone. Due to the modest H_2 requirement for this case, one H_2 plant can serve multiple LNG trains. The primary advantage of this case is its compatibility with existing gas turbines with only minor upgrades.

The full H_2 retrofit and new-build cases achieved CO_2 emissions reductions of 96.6% and 98%, respectively. These cases show how existing and new plants may achieve a high degree of emissions reduction, though both require significant changes to the gas turbines to accommodate high-H_2 fuels. Both cases show how the pre-combustion capture technology[a] provides a pathway to eliminating the carbon footprint of natural gas liquefaction. **GP**

NOTES

[a] Air Products' AP-Blue™ LNG
[b] Air Products' AP-C3MR™

LITERATURE CITED

1. Mallapragada, D. S., E. Reyes-Bastida, F. Roberto, E. M. McElroy, D. Veskovic and I. Laurenzi, "Life cycle greenhouse gas emissions and freshwater consumption of liquefied Marcellus shale gas used for international power generation," *Journal of Cleaner Production,* 2018.
2. Ott, C. M., J. D. Bukowski, R. Shnitser and J. Dunn, "Efficiency, low CAPEX nitrogen removal for natural gas liquefaction," Gastech 2021, Dubai, September 21, 2021.
3. Saunderson, R. P., "End-flash is totally cool," *Hydrocarbon Engineering*, May 2021.
4. Folger, P., "Carbon capture technology assessment: In brief," U.S. Congressional Research Service, R43300, November 5, 2013.
5. Dewing, R., V. White and J. Hicks, "Hydrogen plants with CO_2 capture: Retro-fit options and practical considerations," 15th International Conference on Greenhouse Gas Control Technologies (GHGT-15) Abu Dhabi, UAE, October 5–8, 2020.

DEJAN VESKOVIC joined Air Products in 2011 as part of the Air Products HyCO business, supporting process safety and plant operations. He now works as a Principal Process Engineer in the LNG Process group, where he has designed and developed LNG plants with a range of liquefaction technologies, as well as supported the startup and performance testing of those plants. He received his BS and MS degrees in chemical engineering from Pennsylvania State University and Villanova University, respectively.

JEREMY D. BEARD is a Lead Machinery Engineer in the Process Gases Group at Air Products and Chemicals Inc., supporting both the development and execution of new projects related to the production of blue H_2 through carbon capture, utilization and storage. Mr. Beard has also provided machinery support to other groups within Air Products in the development of projects and process technology in LNG, syngas and other products. He received his BS degree in mechanical engineering from Louisiana State University.

MARK ROBERTS joined Air Products in 1996 and has more than 30 yr of experience developing cryogenic cycles for gas separation and liquefaction in the air separation, hydrocarbon processing and LNG industries. He has more than 40 U.S. and international patents issued in his name, including the patents for the AP-X™ liquefaction process used for the six mega-trains in Qatar, the patent for the AP-N™ liquefaction process deployed in the first offshore project, Petronas FLNG *Satu,* and the patents for the AP-DMR™ dual-mixed refrigerant liquefaction process used on the Coral South FLNG vessel under construction.

DAVE GRAHAM is a Research Engineer at Air Products and Chemicals. He has worked in the areas of H_2 production, air separation, LNG and carbon capture during his 22-yr career. Dr. Graham earned a BSE degree from the University of Michigan, and MS and PhD degrees from the University of Illinois.

Simplified simulation model for an LNG terminal

J. W. CHANG and D. KOLTE, CTCI Corp., Taipei, Taiwan

The natural gas liquids/liquefied natural gas (NGL/LNG) value chain encompasses the production, processing and liquefaction of natural gas (NG) to LNG, its long-distance transportation, and its regasification as it travels from the wellhead to end users. An LNG terminal is an important part of the NGL/LNG value chain and is used at the final stage to unload LNG and distribute NG to users.

Considering that approximately half of the total cost of an LNG terminal project comes from equipment and critical lines, the finalization of the facility's capacity and equipment size verification are critical during the bidding stage. In general, 30–40 operating cases exist in an LNG terminal, so reviewing and checking all operating cases are time-consuming tasks.

This article will focus on the application of a shortcut rigorous simulation model[1] for front-end engineering design (FEED) to ensure the optimum design for cost estimation during the bidding stage to secure competitive advantage.

Operating cases in an LNG terminal. Operating cases in LNG terminals are based on various factors, such as LNG composition [lean/rich/high nitrogen (N_2)], ship unloading status (unloading/holding—no unloading), NG send-out flowrate from the terminal (maximum/minimum/zero), send-out condition (high-pressure, low-pressure), ship loading rate from the LNG terminal (if the ship loading provision is considered), etc.

With the above prescribed conditions, various required operating cases are summarized here:

1. Ship unloading minimum and zero send-out case (lean/rich/high N_2)
2. Ship unloading maximum send-out case (lean/rich/high N_2)
3. Ship unloading normal send-out case (lean/rich/high N_2)
4. No ship unloading (holding) minimum and zero send-out case (lean/rich/high N_2)
5. No ship unloading (holding) maximum send-out case (lean/rich/high N_2)
6. No ship unloading (holding) normal send-out case (lean/rich/high N_2)
7. NG battery limit send-out pressure (high/low).

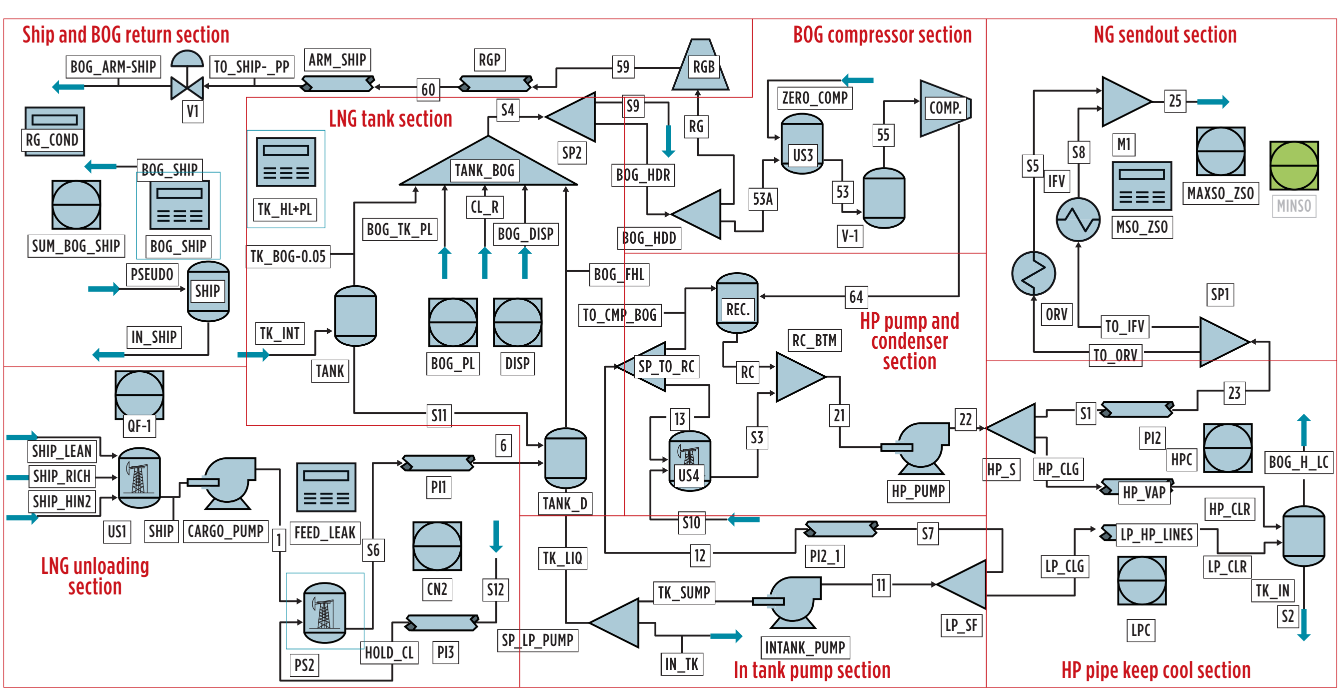


FIG. 1. A typical simplified LNG terminal simulation model.

Given these conditions, some 40–50 operating cases must be evaluated. The results should be reviewed to determine the governing case with a different combination of all equipment and associated connected lines.

For the normal specification above, additional scenarios must be studied if specified in an invitation to bid (ITB) document, such as:

1. Additional operating cases might be needed for two unloading jetties
2. Phase II expansion or any interaction with an existing LNG terminal.

Traditional simulation. To meet the above-mentioned requirement, simulation is the first priority. A traditional simulation model requires various input changes and external calculation assistance for simulating various cases. LNG tank and ship boil-off rates are calculated externally as per client information in the contract and used as an input for the simulation model. Some cases only call for a change in input data; however, for a few operating cases, the structure of simulation model must also be modified.

Input for the general simulation model. The following are the general simulation input/changes required for each simulation case, regardless of unloading/holding mode operation:

- LNG ship and LNG tank unloading/holding mode boil-off rate is calculated externally and used as a simulation input for concerned operating mode cases.
- LNG composition must be input into the simulation model manually based on the considered LNG composition.

Unloading mode. A change required for an unloading mode case is:

- Generally, the unloading mode simulation model is prepared first and used as a basis for preparing further operating cases, including the holding mode.

Holding mode. Considering the unloading mode simulation model is available, the following changes are required for the holding mode simulation model:

- Calculate the ship and LNG tank boil-off based on holding mode operating conditions.
- Stop boil-off gas (BOG) flow from LNG tank to ship.
- Divert the required part of flow from the low-pressure (LP) LNG pump to the unloading line to keep cool purpose.
- Change the controller setting in the simulation model for estimating keep cool flow.
- In circumstances with more than two LNG tanks, ensure proper keep cool LNG flow distribution among LNG tanks based on hydraulics.
- Change LNG tank operating pressure.

For the nine steps mentioned above, > 20 input changes are required besides the actual structure changes in the simulation model.

Simplified simulation model. Critical sections of the LNG terminal simulation model are discussed here to demonstrate the usefulness of automation. By adopting the described approach, users can extensively reduce the input changes required for switching between various operating cases. Users should note that an increase in automation also increases the complexity of the model. With these options, the working time of simulation decreases significantly. Additional automation can be communicated, if needed; however, more automation may increase the convergence issue, and also occasionally increases convergence time. This should be decided case by case. **Note:** General simulation means that no optimization tool or programming are used.

COMMON SIMULATION CHANGES FOR UNLOADING/HOLDING MODE

Ship boil-off. With the help of simple coding, ship boil-off is calculated in simulation by linking actual stream properties. The tutorial code[a] is used inside the calculator block BOG-SHIP in **FIG. 1** (SHIP & BOG RETURN section). Procedure 1 for ship boil-off (for calculator block "BOG_SHIP") is below:

```
$C1: Ship boil-off: VAP vol%/d
    for QMAX ship (0.15 vol%)
$C2: QMAX ship volume
    (266,000 m3)
P(1): Consider LNG stream
    actual density (kg/m3)
P(2): Consider BOG stream
    actual density (kg/m3)
Part 1: R(1)=C(1) × C(2)/100/
    24 × P(1)
Part 2: 10 IF (P(3).GT.10000) then,
    20 R(2)=R(1)/P(2)
    30 else
    40 R(1)=14000 × P(2)
    50 end if
    Return
```

Code description. In Part 1, LNG ship boil-off is calculated based on actual case density. Part 2 is one of the multiple steps for switching unloading mode to holding mode. In this step, ship boil-off was increased to ensure no BOG return to ship; as during holding mode operation, SHIP is not present. C represents the "constant" that can be specified by the user; P represents the "parameter" that can be extracted from the simulation model; and R represents the "calculation results" that are used in further modules.

LNG tank boil-off. LNG tank boil-off is governed by many factors. In general, manual input is required for various cases. However, with the help of a simple program, all boil-off calculations are automated and external calculation assistance is no longer required. Refer to calculator Block "TK_HL + PL" in **FIG. 1** (LNG TANK section). Procedure 2 for LNG tank boil-off (for calculator block "TK_HL + PL"):

```
$C1: Tank height
$C2: Tank diameter
$C3: Tank thickness
    (50+16)/2 = 33 mm
$C4: Tank material density
$C5: Metal heat capacity
$C6: Five sets of tanks
P(1): LNG density (kg/m3)
P(2): Enthalpy of BOG (Kj/kg)
P(3): Enthalpy of LNG (Kj/kg)
P(4): LNG unloading flowrate
    to LNG tank (m3/hr)
Part 1: R(1) = P(2)-P(3)
```

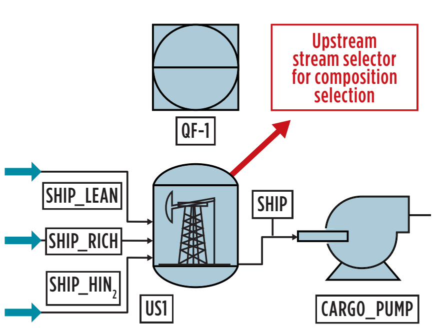


FIG. 2. LNG composition must be input based on various operating cases.

Part 2: R(2) = 3.14159 × C(2) × C(2)/4 × C(1) × P(1) × 0.05/100/24 × R(1)/1E6 × C(6)
Part 3: R(3) = 3.14159 × (C(2) + C(3)) × C(3) × C(4) × (P(4)/ (3.14159 × C(2) × C(2)/4)) × C(5) × 4/1E6
Part 4: R(4) = R(2)/R(1) × 1E6
Part 5: 10 if (P(1).GT.10000), then
 20 R(5) = R(3)/R(1) × 1E6
 30 else
 40 R(5) = 0.001
 50 end if
 Return

Code description. Part 1 details the latent heat calculation; Part 2 addresses heat duty corresponding to LNG tank boil-off; Part 3 covers heat duty that corresponds to incremental cooling of LNG tank plate due to an increase in tank level; Part 4 covers BOG flow—tank boil-off (based on ITB boil-off rate); and in Part 5, this part of the code is used to give a startup value of BOG return to ship to improve convergence time. Heat duty calculated in Parts 2/3 is imparted to a flash that represents the LNG tank.

LNG composition must be input based on various operating cases. In a simulation software[b], "upstream utility" is used to input all LNG compositions once. This utility offers dropdown selection for LNG composition, saving the time to input LNG composition for each case (**FIG. 2**).

Additionally, the whole calculation blocks are created for the send-out section to reduce manual intervention. Refer to the typical simplified LNG terminal simulation model in **FIG. 1** for an overview of the overall simulation model.

Takeaway. The simulation model proposed here is an adaptive and helpful way to conduct quick simulation on the required simulation cases, with a selector button from data specified in the calculator block in lieu of manual input. This allows various operating cases to become more flexible and adaptive. Also, the results can be studied easily and enables better cost estimation during the bidding phase when delivery of proposal timing is challenging. **GP**

NOTES

[a] Fortran Language Quickstart tutorial
[b] AVEVA's PRO-II Simulation software

LITERATURE CITED

[1] Tavares, F. B., T. Mitro, N. Maennling and P. Toledano, Columbia Center on Sustainable Investment, "Manual for the open LNG regasification model," October 2018, online: https://ccsi.columbia.edu/sites/default/files/ content/docs/our%20focus/extractive%20industries/LNG-Import-Model-Manual-CCSI-2018.pdf

JIN-WEN CHANG is a Process Expert within the process department of CTCI Corp. He has 35 yr of experience in the refining, petrochemical and chemical industries. Dr. Chang earned a PhD in chemical engineering from the National Taiwan University of Science and Technology. He is also registered engineer in Taiwan.

DATTATRAY KOLTE is a Senior Process Engineer for CTCI Corp.'s hydrocarbon business operations. He has more than 15 yr of experience in the LNG, refining and petrochemical industries. Mr. Kolte earned a BTech degree in chemical engineering from UDCT–Jalgaon, North Maharashtra University, India.

A competitive study for the behavior of amines in acid gas

N. S. ABBAS, Gulf of Suez Petroleum Co., Cairo, Egypt; **M. Z. A. WAHHAB** and **E. A. ASHOUR,** Minia University, Minia, Egypt; and **T. E. FARRAG,** Port Said University, Port Said, Egypt

Natural gas is the most common fossil fuel. However, because it exists in deep underground reservoirs, it may contain several non-hydrocarbon components [e.g., hydrogen sulfide (H_2S) and carbon dioxide (CO_2)]. These impurities are undesirable compounds, and can cause several technical problems, including corrosion and environmental pollution.

The Gulf of Suez gas plant—located in the Red Sea region of Egypt—has been operated by the Gulf of Suez Petroleum Co. (GUPCO) since 1983. This plant uses diethanolamine (DEA) for the sweetening process. The facility received a significant amount of natural gas containing acid gas that was beyond the plant's design conditions. This resulted in corrosion problems and a shutdown of the amine unit, leading to production losses.

Using proprietary process simulation software[a], plant personnel launched a study aimed at simulating the gas plant's natural gas sweetening process and examining some of the critical amine process factors for each amine type. This article details this process.

TABLE 1. Composition of natural gas feed at the GUPCO gas plant, 2020

Components	Molar flow, $MMft^3d$	Mole fraction
H_2O	0	0
CO_2	0.0588	0.0049
H_2S	0.0072	0.0006
Methane	9.6503	0.8042
Ethane	1.2559	0.1047
Propane	0.5646	0.0471
I-butane	0.0855	0.0071
N-butane	0.178	0.0148
I-pentane	0.0411	0.0034
Nitrogen	0.0806	0.0067
N-hexane	0.0295	0.0025

Aim of the study. Natural gas is a vital component in the world's energy supply. However, despite its importance, there are many challenges when processing natural gas. For example, H_2S is a poisonous and extremely corrosive compound that exists in natural gas. Offshore and onshore plants must remove H_2S, as government regulations make it mandatory to limit H_2S to 4 parts per million (ppm). Over time, various methods for removing H_2S have been successful in plant operations; however, industry is always aiming to optimize the removal process and to create a more efficient sweetening process to minimize costs.[1]

Many different processes have been investigated, but the focus has been to develop a sweetening process by using different kinds of adsorption and absorption methods. When developing gas-sweetening methods, two focus areas are in general use: dry adsorption and wet absorption.[2]

Dry adsorption is usually a non-regenerative method and consists of a filter where H_2S is adsorbed.[3] These filters must be changed out regularly to receive the necessary cleaning. Conversely, wet absorption is usually a regenerative process where a liquid stream reacts with the H_2S in an absorption column to move it from the gas stream to the liquid stream catalyzed by a chemical reaction.[4]

GUPCO's LPG plant treats acid gas with DEA solutions. However, due to corrosion and foaming problems, the plant studied switching from DEA to methyl DEA (MDEA) or another amine. To investigate which natural gas sweetening method worked best, one process was chosen and evaluated thoroughly to evaluate the efficiency of that specific cleaning method. The aim of this study was to determine the most efficient gas sweetening process, using proprietary simulation software[a]. Once the process had been identified, an economic analysis was conducted to ensure that the chosen model is realistic to use.

Process description of the existing gas plant. The operating conditions of rich acid gas feed to the unit include the following:

- Flowrate: 12 $MMft^3d$
- Feed gas temperature: 41°C (105.8°F)
- Feed gas pressure: 49 kg/cm^2
- Concentration of H_2S feed to the contactor: 600 ppm

TABLE 1 provides the 2020 composition of the natural gas feed to the amine unit at the GUPCO gas plant, while the process is shown in **FIG. 1.** The acid gas enters the absorber from the bottom, while the amine solution enters from the top. The amine is made to be reused by passing it through a mixer where it is mixed with a makeup stream of water.

Additional amine solution may be added to the mixer to ensure the optimal composition for gas sweetening. A booster pump increases the pressure to 47 bar, which is the operational pressure of the absorber. A heat exchanger was installed to cool the amine solution to the desired operational temperature of 47.3°C (117.14°F). The sweetened gas from the absorber is sent to the pipe transportation system to be transported to a chilling area for distribution and further processing. The bottom liquid leads to a flash tank, where the stream is heated to flush out most of the hydrocarbon content in the stream. The amine solution then heads to a distillation tower (stripper), where H_2S is removed to the same level as the sweet-

ened gas. The recovered amine is then added with water in the mixer to reuse for the new sweetening process cycle.

Simulation model. To model a realistic gas sweetening process, a specified content of CO_2 must be present. This ensures that the sweetening agent's efficiency and flow are modeled correctly. The normal accepted content of CO_2 in the untreated natural gas should be 4 mol%–5 mol%.[5] In addition, the content of CO_2 in the natural gas stream was 0.0049 mol%. Therefore, water was added to the amine solution. This is done to ensure a larger surface of the liquid phase[6] when a gas stream is added to an absorption column with a relatively high flowrate, the corresponding amount of amine will be relatively low.[7] If water is not added, it will be impossible to ensure a good contact between the gas and liquid, which will cause a less-efficient sweetening process. Thus, water is normally added up to 75 mol% when performing gas sweetening.[8]

The absorption process takes place in a column that contains a certain calculated number of stages. The mass balance used in this analysis was the McCabe-Thiele model. The two inlet streams are the feed gas and the lean amine. The temperature of the amine-water stream should be at least 5°C warmer than the acid gas stream in the absorption column to avoid components of the natural gas condensing into the liquid amine stream. The pressure of the absorption column during operation was around 49 bar, which is normally accepted for the absorption column.[9]

Results of steady-state simulation and process optimization. To evaluate the efficiency of gas sweetening, several treatment cases were modeled in the proprietary simulation software[a]. In all simulations, H_2S content had to be below 4 ppm to match the specifications for clean gas. Nine cases were modeled to investigate the efficiency of tray and packed columns and to evaluate thermodynamic models. The amine treatments used were monoethanolamine (MEA), DEA, diglycolamine (DGA), piperazine (PZ), triethylamine (TEA), diisopropylamine (DIPA) and MDEA. For all cases, the amine packages (for acid gas cleaning) in the modeling were used.[10]

Chemical absorption by amines. Selecting the appropriate solvent for gas sweetening depends on various conditions that can include the ability to remove H_2S and CO_2, the pickup rate of hydrocarbons, and the heat requirement for solvent regeneration, as well as vapor pressure, foaming, selectivity, thermal stability, corrosivity, cost and availability.[11] In this simulation experiment, amine and amine blends were used as sweetening solvent. The chemical reactions for MEA and DGA included the following ("R" refers to amine):

- $RNH_2 + H_2S = RNH_3HS$
- $RNH_2 + H_2O + CO_2 = RNH_3HCO_3$.

Chemical reactions for DEA are:

- $R2NH + H_2S = R2NH_2HS$
- $2R2NH + CO_2 = R2NCOOR2NH_2$.

Chemical reactions for MDEA and DIPA are:

- $R2NCH_3 + H_2S = R2NHCH_3HS$
- $R2NCH_3 + CO_2 + H_2O = R2NHCH_3HCO_3$.

Process optimization. The aim of this optimization work was to study the effects of using various types of amine and amine blends on amine plant performance. Therefore, optimization options were studied by applying various amines at constant concentrations within the simulation to monitor changes in the composition of sour gas streams.

The experiments used 32 wt% of different amine and amine blends in the GUPCO gas sweetening plant by maintaining some operational conditions (e.g, sour gas stream pressure and temperature, and lean amine pressure and temperature). Once the simulation was concluded, the results for H_2S and CO_2 in the sweet gas streams were recorded.

Parametric sensitivity analysis and simulation modeling using MEA. In TABLE 2, the H_2S composition of sweet gas is 4.508e-004 ppm. The recovery for removed H_2S from both the sweet gas and the acid gas (H_2S and CO_2) was 100%. Amine strength was 32 wt% and the amount of regenerated steam required

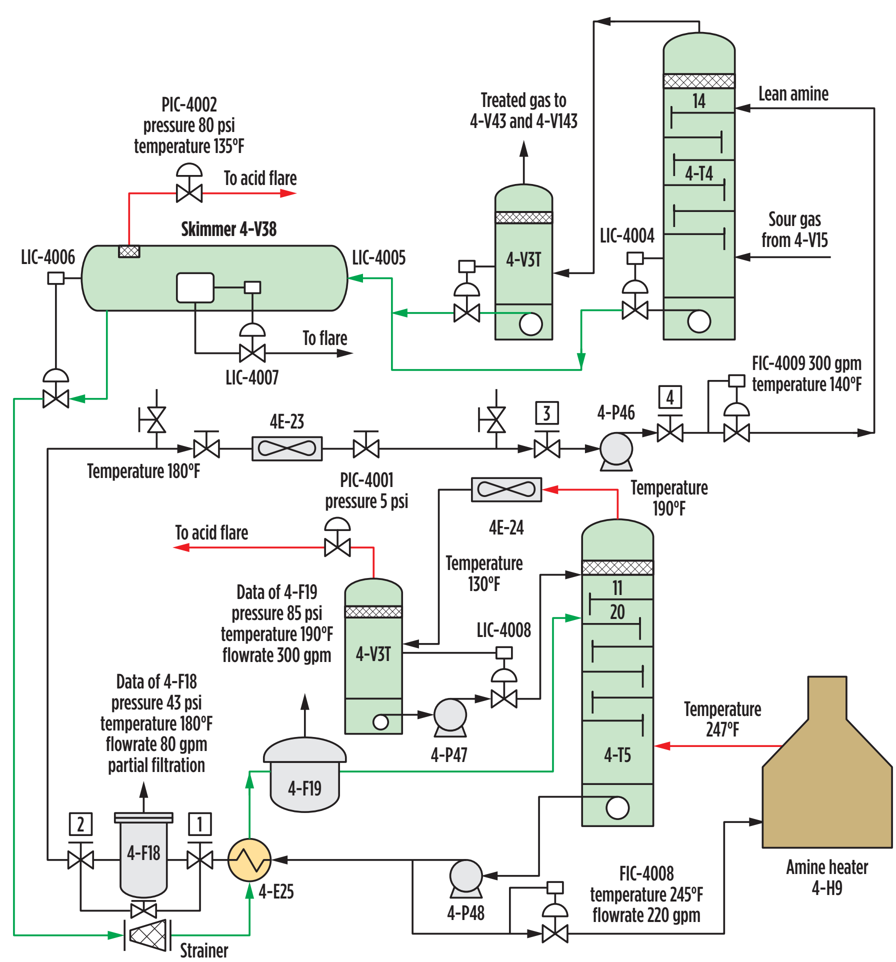


FIG. 1. Process flow diagram of the amine unit at the GUPCO plant.

for the regenerator was 307.4 tpd. Moreover, the total flowrate required was 170 bpd, and the total mass flow required from MEA was 366.5774 kg/hr. The composition of sweet gas by mole fraction for methane was 0.806; for ethane, it was 0.1049. The operating condition for the rich amine stream was 48.03 kg/cm^2 and the temperature was 39.27°C (102.69°F).

Parametic sensitivity analysis and simulation modeling using DEA. Treatment by DEA is shown in **TABLE 3.** The H_2S composition in the sweet gas was 4 ppm. The recovery for removed H_2S was 99.33%, while the recovery of both H_2S and CO_2 was 97.47%. The amine strength was 32 wt%, and the amount of regenerated steam required for the regenerator was 307.4 tpd.

After the trial, it was found that the best flowrate of DEA to meet specifications was 310 bpd, and, to meet the concentration of H_2S in sales gas specifications, the total mass flowrate required from DEA was 683.806 kg/hr. In addition, the concentration (mole fraction) of methane in the sweet gas stream was 0.8057; for ethane, it was 0.1049. The operating conditions of the rich amine included a pressure of 48.03 kg/cm^2 and a temperature of 38.98°C (102.16°F).

Parametic sensitivity analysis and simulation modeling using TEA. Treatment by TEA is shown in **TABLE 4.** H_2S composition in the sweet gas was 0.2037 ppm. The recovery for removed H_2S was 99.97%, while the recovery of both H_2S and CO_2 was 70.29%. The total mass flowrate of TEA to recover this amount was 2,220 kg/hr. The amine strength was 32 wt%, and the amount of regenerated steam required for the regenerator was 307.3 tpd. In addition, the concentration (mole fraction) of methane in the sweet gas stream was 0.8042; for ethane, it was 0.1047. The operating conditions of the rich amine included a pressure of 48.03 kg/cm^2 and a temperature of 39.22°C (102.6°F). The total mass flowrate required to remove acid gas by 70.29% was 1,000 bpd.

Parametic sensitivity analysis and simulation modeling using DGA. Treatment by DGA is shown in **TABLE 5.** H_2S composition in the sweet gas was 8.908e-002 ppm. The recovery for removed H_2S was 99.99%, while the recovery of both H_2S and CO_2 was 99.62%. The amine strength was 32 wt%, and the amount of regenerated steam required for the regenerator was 307.3 tpd. The total mass flowrate required from DGA was 608.73 kg/hr. The concentration (mole fraction) of methane in the sweet gas stream was 0.8058; for ethane, it was 0.1049. The operating conditions of the rich amine included a pressure of 48.03 kg/cm^2 and a temperature of 38.77°C (101.8°F). The total mass flowrate required to remove acid gas by 99.62% was 282 bpd.

TABLE 2. MEA unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.495e+006
Regenerator steam, tpd	307.4
Acid gas loading in regenerator bottom	1.381e-003
Acid gas loading in acid stream	8.537e+014
Acid gas loading in regenerator feed	0.5495
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	161.1
Amine strength, wt%	32
Amine recirculation rate, bpd	170
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft^{3}d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	4.508e-004
H_2S composition in acid gas, ppm	9.603e+004
H_2S composition in regenerator bottom, ppm	0.7344
H_2S composition in regenerator feed, ppm	7,083
Recovery of amine unit for H_2S, %	100
Recovery of amine unit (H_2S and CO_2), %	100

TABLE 3. DEA unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.494e+006
Regenerator steam, tpd	307.4
Acid gas loading in regenerator bottom	2.125e-004
Acid gas loading in acid stream	5.178e+024
Acid gas loading in regenerator feed	0.4927
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	166.6
Amine strength, wt%	32
Amine recirculation rate, bpd	310
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft^{3}d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	4
H_2S composition in acid gas, ppm	9.866e+004
H_2S composition in regenerator bottom, ppm	0.1783
H_2S composition in regenerator feed, ppm	4,027
Recovery of amine unit for H_2S, %	99.33
Recovery of amine unit (H_2S and CO_2), %	97.47

TABLE 4. TEA unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.492e+006
Regenerator steam, tpd	307.3
Acid gas loading in regenerator bottom	3.012e-005
Acid gas loading in acid stream	8.086e+027
Acid gas loading in regenerator feed	0.1553
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	225.1
Amine strength, wt%	32
Amine recirculation rate, bpd	1,000
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft^{3}d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	0.2037
H_2S composition in acid gas, ppm	1.344e+005
H_2S composition in regenerator bottom, ppm	5.208e-007
H_2S composition in regenerator feed, ppm	4,027
Recovery of amine unit for H_2S, %	99.33
Recovery of amine unit (H_2S and CO_2), %	97.47

Parametric sensitivity analysis and simulation modeling using MDEA. Treatment by MDEA is shown in TABLE 6. H_2S composition in the sweet gas was 3.625 ppm. The recovery for removed H_2S was 99.4%, while the recovery of both H_2S and CO_2 was 64.58%. The amine strength was 32 wt%, and the amount of regenerated steam required for the regenerator was 306.3 tpd. The total mass flowrate required from MDEA was 707.48 kg/hr. The concentration (mole fraction) of methane in the sweet gas stream was 0.8046; for ethane, it was 0.1047. The operating conditions of the rich amine included a pressure of 48.03 kg/cm² and a temperature of 38.71°C (101.7°F). The total mass flowrate required to remove acid gas by 64.58% was 324 bpd.

Parametric sensitivity analysis and simulation modeling using DIPA. Treatment by DIPA is shown in TABLE 7. H_2S composition in the sweet gas was 355.6 ppm. The recovery for removed H_2S was 40.7%, while the recovery of both H_2S and CO_2 was 85.12%. The amine strength was 32 wt%, and the amount of regenerated steam required for the regenerator was 307.6 tpd. The total mass flowrate required from MDEA was 774.95 kg/hr. This data showed that the concentration of H_2S and CO_2 were with the trays at the top of the absorber. However, H_2S concentration increased until Tray 6, and then decreased from Tray 5 and so on.

The concentration (mole fraction) of methane in the sweet gas stream was 0.8051; for ethane, it was 0.1048. The operating conditions of the rich amine included a pressure of 48.03 kg/cm² and a temperature of 38.73°C (101.7°F). The total mass flowrate required to remove acid gas by 85.12% was 360 bpd.

Parametric sensitivity analysis and simulation modeling using PZ treatments. Treatments by PZ activators are shown in TABLE 8. H_2S composition in the sweet gas was 1.872e-002 ppm. The recovery for removed H_2S was 100%, while the recovery of both H_2S and CO_2 was also 100%. The amine strength was 32 wt%, and the amount of regenerated steam required for the regenerator was 307.4 tpd. The total mass flowrate required from PZ was 357.3 kg/hr. The concentration (mole fraction) of methane in the sweet gas stream was 0.8059; for ethane, it was 0.1049. The operating conditions of the rich amine included a pressure of 48.03 kg/cm² and a temperature of 39.03°C (102.3°F). The total mass flowrate required to remove acid gas by 100% was 165 bpd.

Parametric sensitivity analysis and simulation modeling using MDEA with sulfolane. Treatments using MDEA with sulfolane are shown in TABLE 9. H_2S composition in the sweet gas was 1.549 ppm. The recovery for removed H_2S was 99.74%, while the recovery of both H_2S

TABLE 5. DGA unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.493e+006
Regenerator steam, tpd	307.3
Acid gas loading in regenerator bottom	1.822e-004
Acid gas loading in acid stream	1.036e+023
Acid gas loading in regenerator feed	0.5655
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	161
Amine strength, wt%	32
Amine recirculation rate, bpd	282
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft³d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	8.908e-002
H_2S composition in acid gas, ppm	9.606e+004
H_2S composition in regenerator bottom, ppm	2.624e-002
H_2S composition in regenerator feed, ppm	4,547
Recovery of amine unit for H_2S, %	99.99
Recovery of amine unit (H_2S and CO_2), %	99.62

TABLE 6. MDEA unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.495e+006
Regenerator steam, tpd	306.3
Acid gas loading in regenerator bottom	3.308e-004
Acid gas loading in acid stream	1.056e+019
Acid gas loading in regenerator feed	0.3579
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	251.7
Amine strength, wt%	32
Amine recirculation rate, bpd	324
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft³d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	3.625
H_2S composition in acid gas, ppm	1.484e+005
H_2S composition in regenerator bottom, ppm	3.012e-004
H_2S composition in regenerator feed, ppm	3,967
Recovery of amine unit for H_2S, %	99.4
Recovery of amine unit (H_2S and CO_2), %	64.58

TABLE 7. DIPA unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.498e+006
Regenerator steam, tpd	307.6
Acid gas loading in regenerator bottom	3.632e-004
Acid gas loading in acid stream	1.24e+023
Acid gas loading in regenerator feed	0.4812
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	189.5
Amine strength, wt%	32
Amine recirculation rate, bpd	360
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft³d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	355.6
H_2S composition in acid gas, ppm	4.5968e+004
H_2S composition in regenerator bottom, ppm	8.939e-005
H_2S composition in regenerator feed, ppm	1,490
Recovery of amine unit for H_2S, %	40.7
Recovery of amine unit (H_2S and CO_2), %	85.12

and CO_2 was 67.97%. The amine strength was 32 wt% (32% for MDEA and 2% for sulfolane), and the amount of regenerated steam required for the regenerator was 307.3 tpd. The total mass flowrate required from MDEA was 789.51 kg/hr and 49.3 kg/hr for sulfolane. The concentration (mole fraction) of methane in the sweet gas stream was 0.8047; for ethane, it was 0.1047. The operating conditions of the rich amine included a pressure of 48.03 kg/cm^2 and a temperature of 38.81°C (101.9°F). The total mass flowrate required to remove acid gas by 67.97% was 360 bpd.

Parametic sensitivity analysis and simulation modeling using MDEA with a PZ mixture. Treatments by the PZ activators with MDEA are shown in **TABLE 10.** H_2S composition in the sweet gas was 5.211-002 ppm. The recovery for removed H_2S was 100%, while the recovery of both H_2S and CO_2 was also 100%. The amine strength was 32 wt% (16% for MDEA and 16% for PZ), and the amount of regenerated steam required for the regenerator was 306.3 tpd. The total mass flowrate required from MDEA with PZ was 435 kg/hr. The concentration (mole fraction) of methane in the sweet gas stream was 0.8058; for ethane, it was 0.1049. The operating conditions of the rich amine included a pressure of 48.03 kg/cm^2 and a temperature of 39.25°C (102.7°F). The total mass flowrate required to remove acid gas by 100% was 200 bpd.

Takeaway. This study used proprietary simulation software[a] to test GUPCO's gas sweetening process design calculations. While the plant receives high amounts of acid gas, this challenge can be solved by changing the type of amine used at the gas sweetening plant. In addition, the simulation study achieved a high rate of acid removal for nearly all amine types and blends, enabling the natural gas to meet pipeline specifications. The optimal solution was 32% MEA with a circulation rate of 170 bpd.

The study also examined several process parameters, including reboiler duty, regenerator steam and acid gas removal. According to the simulation, 32% MEA is the most recommended amine for the process; however, PZ and MDEA at 32 wt% (16% PZ and 16% MDEA) reached full efficiency characteristics. A recirculation rate of 200 bpd was required, though, while the PZ activator needed a recirculation rate of 165 bpd. The cost of PZ chemicals is more expensive than the cost of MEA chemicals, as well. **GP**

NOTE

[a] Aspen HYSYS V8.8

LITERATURE CITED

Complete literature cited available online at www.GasProcessingNews.com.

TABLE 8. PZ unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.493e+006
Regenerator steam, tpd	307.4
Acid gas loading in regenerator bottom	1.93e-003
Acid gas loading in acid stream	7.734e+015
Acid gas loading in regenerator feed	0.7949
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	162.4
Amine strength, wt%	32
Amine recirculation rate, bpd	165
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft^{3}d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	1.872e-002
H_2S composition in acid gas, ppm	9.68e+004
H_2S composition in regenerator bottom, ppm	1.876
H_2S composition in regenerator feed, ppm	7,525
Recovery of amine unit for H_2S, %	100
Recovery of amine unit (H_2S and CO_2), %	100

TABLE 9. MDEA with sulfolane unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.493e+006
Regenerator steam, tpd	307.3
Acid gas loading in regenerator bottom	3.377e-004
Acid gas loading in acid stream	7.242e+018
Acid gas loading in regenerator feed	0.3375
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	232.5
Amine strength, wt%	32
Amine recirculation rate, bpd	360
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft^{3}d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	1.549
H_2S composition in acid gas, ppm	1.376e+005
H_2S composition in regenerator bottom, ppm	3.21e-005
H_2S composition in regenerator feed, ppm	3,650
Recovery of amine unit for H_2S, %	99.74
Recovery of amine unit (H_2S and CO_2), %	67.97

TABLE 10. MDEA with PZ unit dashboard

Component	Value
Reboiler duty, kcal/hr	6.493e+006
Regenerator steam, tpd	306.3
Acid gas loading in regenerator bottom	4.726e-004
Acid gas loading in acid stream	4.354e+016
Acid gas loading in regenerator feed	0.756
Regenerator feed temperature, °C (°F)	104.4 (219.92)
Regenerator reflux ratio (mole basis)	162.4
Amine strength, wt%	32
Amine recirculation rate, bpd	200
Lean amine temperature, °C (°F)	48 (118.4)
Feed gas flowrate, MMft^{3}d	12
H_2S composition in feed gas, ppm	599.6
H_2S composition in sweet gas, ppm	5.211e-002
H_2S composition in acid gas, ppm	9.683e+004
H_2S composition in regenerator bottom, ppm	5.379
H_2S composition in regenerator feed, ppm	6,303
Recovery of amine unit for H_2S, %	99.99
Recovery of amine unit (H_2S and CO_2), %	100

Case study: The science behind amine process column corrosion and remedial solutions

D. SNIJDERS, IGS, Al Khobar, Saudi Arabia

This case study examines how and why corrosion takes place within amine vessels, the corrosion mitigation strategies available, and a comparison of high-velocity thermal spray (HVTS) with a low-velocity alternative.

In 2015, a gas plant in the Middle East identified internal corrosion in two of its amine process columns and contracted the author's company to apply HVTS alloy cladding to protect the base metal. The HVTS cladding was applied, and in 2016, the plant went out to public tender for the completion of a similar work scope on six more columns. The plant decided to use a local thermal spray contractor on a cost basis, which involved using a conventional low-velocity thermal spray process technology.

Amine vessel corrosion mechanisms. To understand the solution to the problem, it is important to first understand the science behind the problem of corrosion—in this case, the amine vessel corrosion mechanisms. In the amine system, most of the corrosion is related to acid gas breakout and the subsequent attack on the metal surface, usually at elevated temperatures. Since corrosion is a chemical reaction, high temperatures always accelerate corrosion activity because reactions occur faster and more aggressively at higher temperatures.

Often, several factors contribute to failure by corrosion. Statistics show that almost 50% of corrosion incidents occur in the hottest part of the plant: the reboiler and the bottom of the amine regenerator. In amine plants, most corrosion is related to hydrogen sulfide (H_2S) or carbon dioxide (CO_2) breakout and the subsequent attack on the metal surfaces, usually in areas of elevated temperature or high pressure drop (leading to condensation).

More potential corrosion mechanisms exist in amine columns. In the lower section of a regenerator, for example, carbon steel surfaces not wetted by the amine solution may be attacked by water vapor and the formation of carbonic acid. Acid gas ratio, choice of amine (DEA, MEA, MDEA, DGA, AGR, etc.), contaminants, two-phase flow, flashing, high velocities, vessel design and insulation are all contributing factors. Pitting, crevice corrosion, flow-enhanced corrosion and general corrosion loss are frequent problems.

In this case, the plant in the Middle East conducted a planned inspection and noticed pitting in its amine process columns that would eventually lead to failure if left untreated and potentially cost the plant millions of dollars to replace the columns.

Several solutions are available to treat the issue of corrosion; while the damage cannot be reversed, if repaired using the correct technique and application, the lifespan and performance of the equipment can be significantly increased.

Corrosion mitigation strategies in amine treatment. Many amine towers have the top dome and upper area of the vessel clad with 316 stainless steel as part of the original design. Process control and changes to the feed typically mean that the amine column will experience corrosion in other areas of the vessel, often directly below the clad section. The nature and form of the corrosion mechanism often manifest themselves as a localized pitting attack with high corrosion rates. That can quickly remove the corrosion allowance and damage the integrity of the unit, requiring intervention and mechanical repair to rebuild the pressure boundary.

Aside from operational considerations, several mitigation or repair strategies have historically been employed when internal shell loss occurs. Some mechanical options include the installation of temporary clamps and plugs, vessel section replacement, and the application of an internal weld overlay. The time required for these, as well as the post-weld heat treatment (PWHT) required to remove heat affected zones (HAZs) and structural support considerations, often demand extended turnaround or shutdown schedules and associated production losses. Temperature and chemical compatibility limitations have prevented the effective use of organic coatings. Due to the nature of the sour feed, loss of containment poses major safety and environmental risks.

The gas plant in the Middle East chose to use an alloy cladding solution to fix the issue of corrosion in its amine process columns. This solution provides the same benefits as a high-nobility weld overlay corrosion barrier, with several additional advantages—including a rapid application speed—without requiring an HAZ or PWHT, or distortion of base

FIG. 1. The 2021 inspection of HVTS in 2015 (applied by the author's company) shows the cladding in excellent condition.

FIG. 2. The 2021 inspection of the six columns coated with a low-velocity thermal spray showed severe corrosion and risk of failure.

materials. In the following section, the results of the application of HVTS on two of the columns vs. the low-velocity thermal spray applied by a local contractor on six of the columns in 2016 are discussed.

The inspections: High-velocity vs. low-velocity thermal sprays. The first inspection of the two columns protected by the author's company's HVTS took place in 2018. No signs of further corrosion were found, and the protective cladding was in excellent, as-applied condition.

An additional inspection by the plant's senior corrosion engineer in November 2021 revealed a staggering contrast between the two columns sprayed with HVTS in 2015 (**FIG. 1**) and the other columns applied with the low-velocity thermal spray in 2016. The HVTS was still in excellent condition, with no signs of cracking or failure. However, the other columns sprayed with the low-velocity thermal spray (shown in **FIG. 2**) indicated extreme signs of breakdown of the applied cladding and corrosion of the substrate, with the wall thickness loss threatening to exceed its maximum corrosion allowance.

If left, the continued corrosion could result in an unplanned shutdown, potentially costing millions of dollars in lost revenue. In the best-case scenario, the column would need to be stripped of the remaining cladding, undergo mechanical repairs (which is both costly and time consuming), and then be protected with a long-lasting HVTS.

The science behind HVTS, material modification and application technology. This scenario is a common occurrence at many oil and gas, refining and petrochemical sites. Although the initial cost of low-velocity thermal spray is often much lower, the long-term expense far outweighs the short-term savings. To understand why the low-velocity system failed, the science behind various types of materials and application methodology must be explored.

Standard welding alloys are unsuitable materials for thermal spray application in critical internal process equipment, as the cladding particles form oxides on their surface in flight, creating a fundamental weakness in the microstructure and permeability. The author's company develops alloys designed and modified for high-performance thermal spray application.

The author's company's HVTS conveyance technologies atomize the bespoke wire feedstock in a supersonic gas stream, producing a cladding that consists of flat and tightly packed microsized particles. Conventionally available thermal spray application systems cannot produce equivalent particle sizes—they have a more open microstructure and do not impart sufficient energy to flatten the particles that are inherently highly stressed. These shortcomings lead to premature failure due to permeation from microstructural weaknesses and cracking of the coating.

Takeaway. The results seen at this gas plant confirm the performance difference between applying an HVTS vs. a conventional thermal spray, eliminating the corrosion process. Visually seeing the difference between the columns in this case study sprayed in 2015 and the columns sprayed with a low-velocity thermal spray is reassuring to owner-operators. The system and application process have been successful in preventing any further degradation or wastage and resulted in significant savings for the client. **GP**

DENNIS SNIJDERS is the Senior Regional Director, Middle East and North Africa, for IGS. He has worked for more than 20 yr in the corrosion industry, 15 yr of which have been spent in the Middle East. In his present position, he is responsible for managing the Middle East and North Africa regions for the entire surface technology solution portfolio of IGS.

Versatile and capable: Advantages of regenerative turbine pumps in LPG applications

S. BASCLAIN, Ebsray, Sydney, Australia

Autogas (also referred to as LPG) has become a popular fuel source around the globe, evidenced by its steady increase in consumption in recent years. In 2019, global LPG consumption reached 27.1 MMt, with 27.8 MM LPG vehicles in use around the world, according to the World LPG Association.

This fuel source is a mixture of propane and butane. Its composition allows it to be stored and transported as a liquid and burned as a gas. With an abundance of LPG vehicles around the world, dispensing installations are needed to meet increasing demand.

However, LPG is not the easiest to handle or process. Pumps designed to handle LPG must be able to handle low viscosities—sometimes as low as 0.1 centipoise (10 times thinner than water)—and vapor pressure well above normal atmospheric pressure. Challenges arise under these conditions as many pumping technologies have trouble with slippage or keeping the substance contained.

Cavitation is another issue that comes with pumping volatile liquids. If the pump's inlet pressure falls below the liquid's vapor pressure, then vapor bubbles will form in the liquid. These bubbles will travel through the pumping chamber and, as the pressure increases, implode and cause cavitation, which can damage the pumping hardware.

Regenerative turbine pumps function as a superb solution to the demands of processing LPG. They can maintain high pressures while handling low-viscosities fluids and are not susceptible to cavitation damage.

This article will examine the capabilities of regenerative turbine pumps and why they work well vs. other pumping technologies when handling LPG and similar substances.

A closer look at regenerative turbine pumps. Though considered rotodynamic, regenerative turbine pumps have close parallels to positive displacement (PD) pumps. Regenerative turbine pumps provide multi-stage performance from a single-stage impeller designed to optimize performance. This enables the pump to create high differential pressures at low flowrates.

Regenerative turbine pumps function using a rotating, non-contact, free-wheeling impeller disc that has around 60 small cells on its periphery. When the liquid enters the suction port, the impeller obtains it and accelerates around the narrow hydraulic channel surrounding the cells. Kinetic energy carries the liquid rapidly around the channel. This motion within one revolution builds energy and pressure (FIG. 1).

The pump technology name stems from its function—the continuous regeneration of small liquid cells creates the differential pressure capability of the pump. Other names for this technology include peripheral pumps, centrifugal regenerative pumps and regenerative pumps, among many others. Regardless of the name, this technology is categorized in the rotodynamic family of pumps.

Regenerative turbine pumps have the versatility to transfer liquids at high pressure and low flow, while also processing entrained vapors or liquids at or near their boiling point. These conditions limit most pump technologies, causing performance issues, cavitation, vibration and noise. However, regenerative turbine pumps do not falter from those conditions.

Specifically, regenerative turbine pumps can handle viscosities of 0.1 centistokes (cSt)–50 cSt, with differential pressures up to 300 psi (20 bar) and a maximum allowable working pressure of up to 493 psi (34 bar) to enable handling liquids with high vapor pressures.

Typical regenerative turbine pumps generate flowrates up to 52.8 gpm (200 l/min), though some variations can operate with flowrates above that amount (FIG. 2). Some newer iterations of this technology can reach peak flowrates as high as—and potentially higher than—158.5 gpm (600 Ll/min).

This pump technology thrives because of its design. The spiral motion, along with its speed, diminishes the chances for cavi-

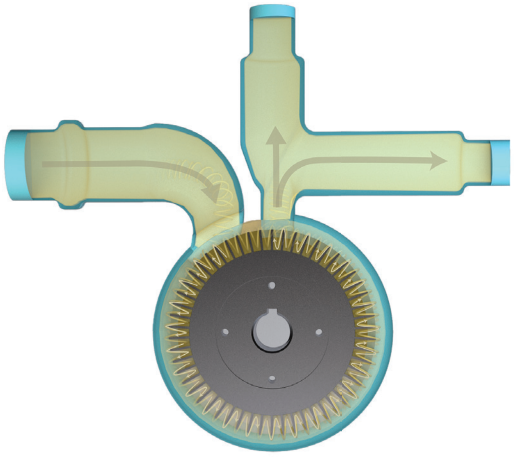


FIG. 1. When liquid enters the suction port of the turbine pump, it is picked up by the impeller and instantly accelerated around in the narrow hydraulic channel (casing volute) surrounding the cells. The spiraling of the liquid at sonic speeds within the one revolution incrementally builds energy and pressure.

tation and pulsation by smoothing the fluid and collapsing the vapor bubbles immediately when they form. Vibration and noise problems also do not occur in most situations since the pump operates with a smooth flow and a hydraulically balanced design.

LPG is not the only application where regenerative turbine pumps excel. This technology functions optimally in applications known for having low-viscosity fluids, such as aerosols and refrigerants. Other applications include ammonia, vaporizer feed and cylinder filling, as well as boiler feedwater.

Matching up with side-channel pumps. Regenerative turbine pumps have similarities to PD pump technologies, such as side-channel pumps. One similarity is both pumps can self-prime and perform optimally under poor suction conditions. The differences come from the size of the pump and ease of maintenance.

Side-channel pumps are designed to have a larger footprint, primarily due to having multiple pumping stages. When considering an LPG installation, a side-channel pump might require 4–8 stages to meet the duty parameters. With that many stages, side-channel pumps and their large footprint become more complex to accommodate the installation's demands.

Regenerative turbine pumps, which feature a single stage, can match the performance of the side-channel pump, while also functioning at two-pole speeds. Side-channel pumps typically operate at four-pole speeds.

With a larger size and more complexity, side-channel pumps require more components to properly function. More components mean more wear parts, all of which are prone to eventual failure. When that happens, those parts must be replaced, adding to the pump's maintenance and total ownership cost.

Regenerative turbine pumps do not suffer from a large footprint or a complex design. Instead, regenerative turbine pumps have a compact footprint and up to 25 components, making it easier for operators to maintain over time. Spending less time on maintenance and components saves time and money. Also, the less-complex design means operators do not need a veteran engineer to maintain them, allowing any technically savvy associate with moderate experience to keep the technology running.

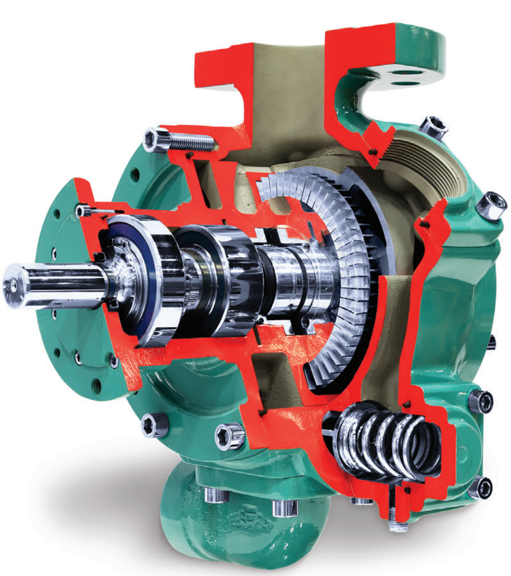


FIG. 2. Typical regenerative turbine pumps generate flow rates up to 52.8 gpm (200 l/min), with some variations capable of even higher flow rates, such as 158.5 gpm (600 l/min).

Stacking up with other PD pumps. Sliding vane pumps also match up well with regenerative turbine pumps and even have some advantages. The primary advantage comes from the technology having a higher hydraulic efficiency and better effectiveness during priming.

Despite regenerative turbine pumps not sharing these traits, they have others that make the technology effective in similar applications. Without as many moving parts as sliding vane pumps, regenerative turbine pumps can operate continuously without many drawbacks.

Operators do not have to worry about taking regenerative turbine pumps offline to replace wear parts on a frequent basis. Similar pump technologies have a variety of wear parts, each with different service and lifecycles. Without as many moving parts, operators of regenerative turbine pumps spend less time shutting down the pump for scheduled maintenance and do not have to stock as many components for eventual replacement.

Regenerative turbine pumps have two components subject to wear: the mechanical seal and (occasionally) the impeller. During scheduled maintenance, operators do not have to take them out of service. Regenerative turbine pumps have parts that can be replaced within an hour without removing the pump from the pipework or even disconnecting the motor.

Performance wise, regenerative turbine pumps can operate continuously without pulsation and cavitation. With the ability to operate continuously, engineers do not have to worry about frequent stops or the aforementioned detriments.

Takeaway. When it comes to challenging applications, especially those with poor suction conditions, low-viscosity liquids or fluids near their boiling point, operators have a wealth of pumping choices. However, if they want a technology with versatility at its core, regenerative turbine pumps stand out among the rest.

These pumps can handle a vast range of liquids under varying conditions in multiple applications. The common deterrents—cavitation, vibration and entrained vapor—that challenge most pumping technologies do not have the same effect on regenerative turbine pumps. Operators can expect longevity from this technology, as well as long intervals between maintenance.

When maintenance rolls around, less-experienced operators are as capable as veterans when servicing this technology. Operators are not required to replace several components or remove the pump for service for several hours. In most cases, the pumps can be repaired or rebuilt without removing them from the pipework.

The value of regenerative turbine pumps in LPG applications is clear: they have the versatility to process it optimally and effectively. As the technology continues to improve in the future, it will become a more common choice among operators in a wide range of applications. **GP**

STEPHEN BASCLAIN is the Business Development Manager for Ebsray. Ebsray is a leader in the design and manufacture of regenerative turbine and positive DP technologies, including sliding vane, gear and lobe pumps.

Safe design and operation of renewable natural gas facilities

M. YOUNG, Baker Engineering and Risk Consultants Inc., Chicago, Illinois

The U.S. produces almost 70 MMtpy of waste[1]; over time, this waste generates biogas with the help of bacteria. Biogas is a mixture of methane (CH_4), carbon dioxide (CO_2) and other trace gases and is normally released into the atmosphere. However, waste collection plants are now exploring ways to collect the biogas and refine it into pure CH_4, also known as renewable natural gas (RNG). RNG can be used for injections into existing natural gas pipelines or used as fuel for RNG-dependent transportation vehicles. In recent years, the number of RNG plants has grown significantly within the U.S.

Despite the many benefits and solutions that RNG provides, hazards remain; because RNG is a relatively new technology, these hazards are not as commonly known. Most RNG plants are small and not covered under Occupational Safety and Health Administration (OSHA) PSM 1910.119, which is required if a facility contains > 5 t of flammable gas or liquid. However, when dealing with flammable and toxic materials, it is important to recognize the potential hazards and/or risks associated with the process regardless of formal requirements.

Examples of previous incidents and studies conducted by the author's company illustrate the significant hazards that are present in RNG plants:

1. An explosion in the anaerobic digester (AD)
2. A rupture of the animal waste AD from vacuum or pressure build-up inside the AD
3. A leak of toxic and flammable material into the atmosphere
4. An explosion in an enclosed compressor building.

This article discusses each of the events described above, highlights the potential hazards and provides examples of potential safeguards and/or mitigations that can be employed to reduce injury or fatality.

The RNG process. The process to produce RNG starts as waste (e.g., animal, landfill or water waste). The feedstock undergoes anaerobic digestion (**FIG. 1**), where organic material is decomposed into biogas.[1,2] Depending on the feed, the digestion can take place in different storage containers called an AD—usually, landfill waste underground, water waste in covered lagoons, and animal waste in non-pressurized tanks. After several weeks, the generated biogas can be collected. Water, hydrogen sulfide (H_2S), CO_2 and other trace chemicals are removed from the biogas through various separation steps to produce the high-quality CH_4 product stream that can be compressed and sent to end users.

AD explosion. The AD's sole purpose is to produce and collect the flammable biogas, and operating an AD for waste poses flammable, explosive and toxic hazards. By design, air is normally not present in the AD. However, with poor maintenance or operation, a leak in the equipment could allow air to enter the AD equipment—with ignition, an explosion can occur, similar to the 2018 in-ground AD tank explosion[3,4] at the Calumet Water Reclamation Plant in Chicago, Illinois, which caused 10 injuries. An investigation concluded that the manhole sealing the in-ground tanks was leaking flammable gas while contractors were conducting hot work nearby. Measures to eliminate nearby ignition sources and to detect and fix the leak could have prevented the explosion in the AD.

H_2S is also produced as a byproduct in the AD and poses toxic hazards that are

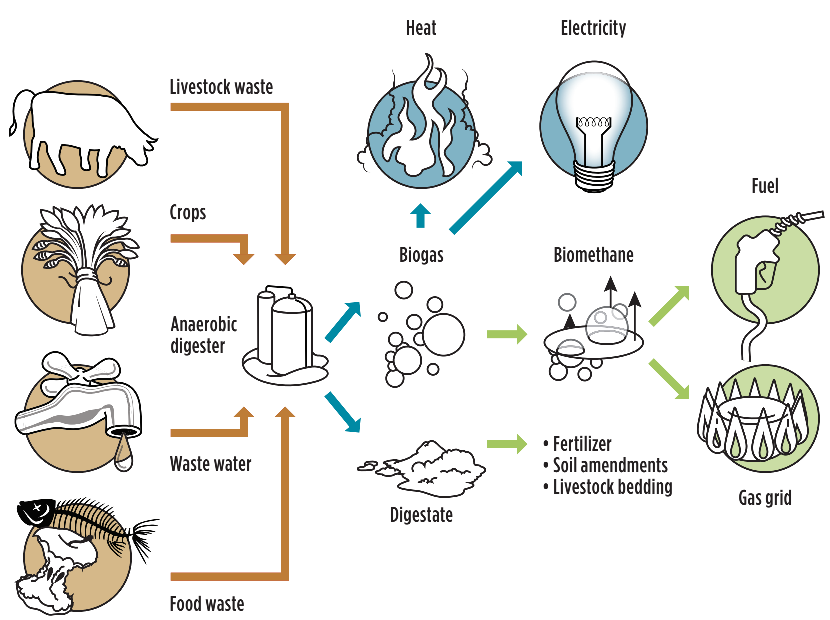


FIG. 1. Process flow for anaerobic digestion (AD).[1]

discussed further in this paper. It is ideal to reduce the amount of H_2S present in the AD to minimize the toxic hazard and the potential corrosion of equipment. A common practice to reduce the production of H_2S is micro-aeration, which is to bleed air [0.3%–3% oxygen (O_2)] into the headspace of the AD.[5] Sulfur-oxidizing bacteria (SOB) are already present in the AD and can easily convert the H_2S and the O_2 into elemental sulfur and water vapor, so this method is cheap and preferred by most facilities operating an AD. While this is an effective method to reduce the toxic hazard of H_2S, it increases the risk of creating an explosive atmosphere and resulting in an explosion within the AD.

Air is added to the AD in a controlled manner to ensure it remains below the lower flammability limit; if the method of adding air fails, the AD can quickly contain an explosive mixture. It is recommended to consider designing a mitigation system to reduce the risk of an AD explosion when incorporating air into the AD with:

- Lower explosive limit (LEL) detection of the AD and alarms in the operator room
- An emergency shutdown protocol to include safety actions to activate appropriate mitigations, such as shutdowns of electrical equipment (ignition sources) or isolation of the AD from oxygen sources
- Frequent testing intervals of mitigation equipment (transmitters, isolation valves, etc.)
- Proper maintenance of mitigation equipment.

Safeguards and mitigations include:

- Preventing the creation of a flammable/explosive environment in the AD:
    - Routine inspection and repair to eliminate potential sources for an influx of air into the AD; this not only eliminates the possibility of a flammable/explosive environment but is also operationally beneficial for the AD.
    - Automatic shutoff of downstream blowers to prevent reverse flow of air into the AD, if applicable.
    - Safety controls to prevent the over bleeding of air or oxygen into the AD, if applicable.
- Reducing ignition potential by determining hazardous area classification (HAC) surrounding the AD and using appropriately classified equipment, especially if using an electrically-driven mixer for the AD.

FIG. 2. Vacuum causing the buckling of AD tank.[7]

FIG. 3. Corrosion causing material to leak from the AD.[9]

AD rupture. Another concern is any loss of biogas containment. The released gases pose flammable and toxic hazards to nearby onsite personnel. If ignited, the flammable gas can lead to a flash or jet fire, while the concentration of H_2S gas can be lethal to personnel close to the AD even without ignition. In cases where chicken manure is used, ammonia (NH_3) is produced as a byproduct,[6] which is a potential toxic hazard even at fairly low concentrations.

Specifically for animal waste, ADs are constructed with floating roofs to account for the growing volume of vapors. However, if not properly monitored, the pressure inside the vessel can increase rapidly to cause the roof to "pop off," or the vessel to rupture. Based on three past events in 2012 and 2013 reported by the UK's Environment Agency,[7] pressure relief valves (PRVs) failed to mitigate overpressures in the AD, causing the rupture of the process vessel and the release of flammable and toxic material into the atmosphere. Overfeeding the AD and poor temperature control can cause increased foaming and liquid level in the vessel. If high enough, this level can block the PRVs and prevent them from operating on demand. Alternatively, the pressure inside the vessel can drop below design pressure. Underfeeding the AD will cause the vessel to vacuum (**FIG. 2**), buckle and release biogas into the atmosphere.[8,9]

Safeguards and mitigations include:

- Properly inspect PRVs
- Closely monitor pressure, level and temperature operations of the AD
- Use process hazard analysis (PHA) studies to determine the appropriate level of safety interlocks needed to prevent over-pressure and vacuum conditions in the AD
- Perform safety instrumented system (SIS) and safety integrity level (SIL) evaluations to ensure safety interlocks meet the desired level of reliability.

Leaks from the AD. Leaks are expected to pose smaller hazards than full-bore ruptures. However, it is more likely in normal operating conditions for smaller leaks to occur with corrosion playing a major factor, as seen in **FIG. 3.** Corrosion of the AD[5] occurs when H_2S created in the process combines with moisture from the AD process to produce sulfuric acid (H_2SO_4), eventually causing releases of biogas but at a lower and more sustainable rate than the rupture of the tank (**FIG. 4**). With smaller rates of release, personnel may experience low doses of toxic gas (H_2S or NH_3), which may be noticed by the pungent odor, especially with H_2S.

The presence of odor can indicate a leak and should be mitigated to prevent additional hazards from occurring.[10]

However, if the odor goes away, it should not be assumed that the hazard has been resolved and is no longer present. It is typical that personnel can become accustomed to the odor or, at higher doses, lose their sense of smell. To prevent leaks, it is essential to ensure material compatibility with all feeds, products and byproducts that may come into contact with the equipment and to inspect and maintain the AD properly.

Safeguards and mitigations include:

- Install H_2S and/or flammable gas detectors near the AD to detect potential leaks
- Conduct routine inspection and repair for corrosion on the AD
- Implement a hot works permit and management of change (MoC) procedures to ensure safe practices are followed to minimize the potential hazards associated with the maintenance
  - Create and follow proper purging protocols of the AD with nitrogen gas (N_2) rather than air to remove pockets of biogas and eliminate an O_2 source prior to hot work.
- Instruct personnel to wear H_2S monitors properly while working in designated areas near the AD and/or other potentially H_2S-rich equipment.

RNG compressor building explosion. At this point in the process, the biogas has been refined into RNG and can be compressed for transport to end users. However, releases from compressors can create a large flammable cloud and pose potential flammable or explosion hazards resulting in potential personnel safety risks. Compressors are typically housed in buildings to protect against the weather; however, if a leak were to occur, the high-pressure gas can quickly accumulate inside the confined structure, resulting in a confined vapor cloud explosion. Unlike the AD explosion, air and ignition sources are typically present in the building.

FIG. 4. Release of biogas from an AD.[7]

FIG. 5. Compressor building explosion in Springville, Pennsylvania.[10]

A small failure of the compressor equipment can quickly provide the flammable concentration required for the explosion.

Several incidents have occurred at natural gas compressor stations, which are comparable to RNG compressors. For example, in Springville, Pennsylvania in 2012,[10] a valve was left open and the compressors were started up after maintenance. Within minutes, the building filled with natural gas and exploded (**FIG. 5**). The gas detection system isolated the feed and discharge lines from the compressors and alerted personnel to evacuate. No injuries were reported and only one of the six compressors was severely damaged, due to the mitigation controls and alarms in place at the facility.

Safeguards and mitigations include:

- Consider the use of a weatherized compressor so it can be outdoors or under a roof structure but not within walls
- If you choose to have compressors indoors, these additional mitigation options should be reviewed and considered for implementation to meet desired risk tolerance:
    - Provide general building ventilation to mitigate small releases of RNG
    - Provide adequate gas detector coverage to accompany the automatic shutoff of feed gas and shutdown of compressors
    - Provide automatic shutoff of RNG feed to compressors as well as automatic shutdown of compressors on gas detection to limit the inventory available for release into the building
    - Design processes with equipment that meet the standards for a Class 1, Div. 1 hazardous area to mitigate the risk of igniting a potential cloud of RNG
    - Design deflagration venting panels on the building to mitigate the severity of an explosion
- Relocate occupied spaces such as control rooms, maintenance offices or other occupancies to separate buildings away from the compressor structure to reduce population vulnerability.

Recommended assessments. Examples of the type of studies that are helpful in the understanding of hazards and risk from the operation of RNG plants include:

- A facility siting study (FSS) and quantitative risk assessment (QRA): These studies are used to model and assess hazards associated with operation of the facility and evaluate the associated risk of those hazards to personnel onsite. An FSS/QRA can assess the risks to personnel associated with site operations, such as AD explosion, over-pressure rupture, biogas release and compressor confined-vapor cloud explosion.
- A hazardous area classification (HAC) study is used to define hazardous areas for electrical classification inside of the compressor building and near the AD tanks.
- A pressure relief valve (PRV) dispersion study is useful to accompany an HAC study because intentional releases from a PRV can become ignition sources to the internal explosion of ADs, based on historical incidents.
- Ventilation design studies are performed to determine the minimum required air exchange rate to ensure adequate ventilation to mitigate confined vapor cloud explosion hazards posed by small releases.
- A fire and gas (F&G) detection study evaluates whether a particular hazardous gas release and/or fire radiation can be detected considering the device location in relation to the hazardous release scenario. Adequate detection is pertinent in the automation of safeguards to mitigate hazards as well as to prevent spurious trips.
- SIS/SIL evaluations are used to verify the safeguards in place and sufficient testing intervals to achieve SIL target.
- Based on the National Fire Protection Agency (NFPA) 68 standard, perform a vent panel study to design the potential mitigation for overpressures from a confined vapor cloud explosion. **GP**

LITERATURE CITED

1 Tanikawa, S., "Fact sheet | Biogas: Converting waste to energy," Environmental and Energy Study Institute (EESI), October 2017, online: https://www.eesi.org/papers/view/fact-sheet-biogasconverting-waste-to-energy

2 U.S. Environmental Protection Agency (EPA), "An overview of renewable natural gas from biogas," EPA 456-R-20-001, July 2020.

3 Wiss, J., "Calumet water reclamation plant—Investigation of explosion and structural collapse in gravity belt thickener room," Elstner Associates Inc., prepared for Metropolitan Water Reclamation District of Greater Chicago, November 16, 2018.

4 Walsh, T., "Calumet water reclamation plant explosion and building collapse: Lessons learned," *Fire Engineering*, June 2020, online: https://www.fireengineering.com/industrial-firefighting/calumet-water-reclamation-plant-explosion-and-building-collapse-lessons-learned/

5 Hines, M., *et al.*, "Microaeration for hydrogen sulfide removal in biogas," University of Maryland, 2019, online: https://projects.sare.org/wp-content/uploads/In-situ-microaeration-Fact-sheet.pdf

6 Krakat, N., *et al.*, "Methods of ammonia removal in anaerobic digestion: A review," *Water Science & Technology*, IWA Publishing, July 2017, online: https://iwaponline.com/wst/article/76/8/1925/19195/Methods-of-ammonia-removal-in-anaerobic-digestion

7 "A review of environmental incidents at anaerobic digestion (AD) plants and associated sites between 2010 and 2018," Environmental Agency 2018, September 2019.

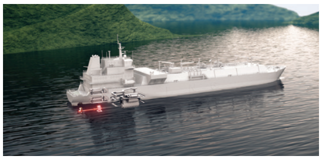
8 "Sludge power plant collapses in Shropshire," *Shropshire Star,* May 2014, online: https://www.shropshirestar.com/news/2014/05/30/shropshire-sludge-power-plant-collapses/

9 Wilkinson, H., "Sludge tank and anaerobic digester corrosion maintenance," Belzona, November 2018, online: https://blog.belzona.com/tank-anaerobic-digester-corrosion/

10 Tanfani, J., "Northern Pennsylvania gas explosion was out of regulatory reach," *The Philadelphia Inquirer,* April 2012, online: https://www.inquirer.com/philly/business/20120408_Northern_Pennsylvania_gas_explosion_was_out_of_regulatory_reach.html

MEGAN YOUNG is a Project I Consultant in the Chicago Process Safety Group, specializing in consequence and risk modeling. Ms. Young has worked extensively on these projects with clients in the chemical processing industry with a focus on risk quantification, refinement and mitigation of blast, toxic, flammable and fire hazards.

ABB releases new shaft generator technology



The new shaft generator from ABB enables ship operators to benefit from the advantages of permanent magnet technology. The generator allows easy and flexible installation to bulk carriers, container carriers and LNG tankers.

The AMZ 1400 permanent magnet shaft generator is optimized for converter control and enables better efficiency than either induction or electrically excited synchronous machines at both full and partial loading. This helps to cut fuel costs and significantly reduce CO_2 emissions.

Its compact size shrinks the footprint of the equipment package by around 20% compared to a conventional synchronous or induction-type shaft generator. In addition, the weight is reduced by approximately 30%. The unit is also simpler to install at the shipyard, as the intermediate propulsion shaft is mounted to the generator and then lifted into the vessel for alignment.

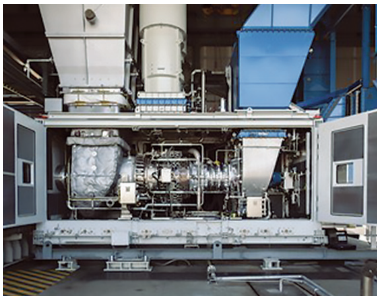
Servomex delivers oxygen analyzer to PTA facility

Servomex has delivered its SERVOTOUGH OxyExact 2200 to INEOS Aromatics for a $70-MM upgrade project at its facilities in Indonesia. In addition, Servomex installed two vent gas analyzer systems. The project is aimed at reducing emissions and increasing capacity.

The SERVOTOUGH OxyExact 2200 is a high-specification oxygen analyzer that monitors process applications such as feedstock purity testing. In addition, it cleans up hydrogen (H_2) and chlorine production, solvent recovery and oxidation control processes, ethylene oxide and ethylene dichloride production.

The technology uses a three-enclosure system that facilitates the sampling and measuring of any flammable gas for 0%–100% oxygen, with no need for pre-sample drying. Up to six analyzers can be connected to a single control unit.

Baker Hughes to provide hydrogen-ready turbo-compression technology for pipeline



Baker Hughes has won a contract from TERNA to supply gas turbines and compressors able to run on a blend of natural gas and hydrogen (H_2) for a new compression station of the Greek Natural Gas Transmission System. The compression station will serve the domestic gas supply in Greece.

Baker Hughes will provide three compression trains for three NovaLT12 H_2-ready gas turbines and three pipeline compressors. The technology has been designed to support the compression station with the capability to transport up to 10% H_2. The station is expected to enter operation in 2024 and directly supports the EU's H_2 strategy goals to accelerate the development of clean H_2 and ensure its role as a cornerstone of a climate-neutral energy system by 2050.

Tellurian's Driftwood LNG begins construction

Tellurian has begun Phase 1 of the Driftwood LNG terminal, an LNG export facility located in Louisiana, U.S., that will have an approximate liquefaction export capability of 27.6 MMtpy.

Once the demolition, civil site preparation and construction of critical foundations are completed, Baker Hughes will manufacture two of the natural gas turbines required to complete Phase 1.

Phase 1 will include two LNG plants with export capacity of 11 MMtpy. Driftwood has obtained all significant permits needed for construction and operation, progressed detailed engineering to about 30% completion and finalized the purchase and lease of approximately 1,200 acres of land.

Drager releases gas detection technology

Drager's X-pid 9500 combines two measuring modes to detect volatile organic substances, such as benzene, at extremely low temperatures in as little as 30 sec. The product is an upgraded version of the X-pid 8500, released in 2017.

The X-pid 9500 includes a search measuring mode, enabling continuous monitoring for PID-sensitive compounds. The user can then utilize the analysis mode for a more in-depth examination.

More than 40 target substances can be measured, including hexane, isobutylene and xylene. Additional substances to target can be added to the database by performing a local qualification, using a known concentration of the target compound.

The search mode uses broadband measuring to determine the total concentration of detectable volatile hydrocarbons in the ambient air. A smartphone can control the gas detection unit with a downloadable application, which also evaluates the measurement data.

LNG Canada receives first mega-module

Canada has received its first significant facility module at the LNG Canada site in Kitimat, B.C.

The 35-m, 4,618-metric t module will be the entry point for natural gas delivered to the LNG Canada site via the new Coastal GasLink pipeline. This module follows several other large pieces of LNG infrastructure recently installed at the site.

Burckhardt Compression, PTC collaborate to develop remote service solutions

Burckhardt Compression is working with PTC to develop remote service solutions for its customers. In addition, the companies are working together to continue developing a type of metaverse augmented reality (AR) solution for service on reciprocating compressors.

Under the name UP! Solutions, Burckhardt Compression offers digital solutions that enable compressors to be monitored in real time and serviced remotely in cooperation with the customer onsite. For this purpose, the units are equipped with sensors that constantly collect data, such as vibrations and temperature, which are then analyzed and used to detect anomalies or schedule service work.

Burckhardt Compression relies heavily on Vuforia, the AR solution from PTC, which observes an expert during a process, compiles instructions for teaching and guiding less-experienced colleagues and then monitors the learning phase.

RINA approves the first hydrogen-powered MR tanker

RINA has received approval in principle for Swedish designer FKAB Marine Design's hydrogen- (H_2-) powered MR tanker. The propulsion is based on combining the ship's fuel source, LNG, with steam in a gas reformer to separate LNG molecules into H_2 and carbon dioxide (CO_2).

The H_2 is then directly utilized to fuel the internal combustion engines and fuel cells. In theory, the ship can meet total decarbonization targets by either running entirely on H_2 or producing all the required power by fuel cells.

Veolia, DuPont convert utility infrastructure

Veolia and DuPont have converted the DuPont Spruance manufacturing facility in Virginia, U.S., from coal to natural gas to efficiently produce steam and electricity. The facility will assist with optimizing energy yields while achieving a GHG emissions reduction of > 220,000 tpy.

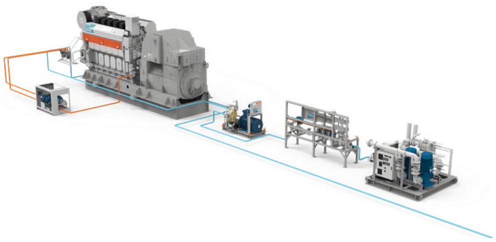
According to project terms, Veolia was to acquire, upgrade, operate and maintain the site's utility infrastructure to improve efficiencies and reliability. The agreement included repurposing a 200-MW cogeneration facility from power production to the production of 500,000 lb/hr of steam, converting from coal to natural gas, and upgrading chilled water capacity, water networks and compressors.

Cynalytica delivers the first deployment of SerialGuard cybersecurity solution for gas pipeline operations

Cynalytica's SerialGuard is an entirely passive serial packet sniffer that allows secure visibility within vulnerable legacy networks designed for industrial control systems (ICSs). Legacy serial communications account for 30%–60% of all critical U.S. infrastructure.

Many ICS security providers only monitor TCP/IP and Ethernet communications, leaving a significant blind spot for gas pipelines. The SerialGuard sensor monitors Level 1 and Level 0 serial communications between field devices and controllers.

Wärtsilä to deliver the first dedicated methanol fuel supply system



Wärtsilä has developed MethanolPac, a fuel supply system specifically for methanol. In conjunction with the newly released Wärtsilä 32 methanol engine, the company can deliver methanol-capable fuel and power systems across a wide range of vessel segments. When methanol is produced from renewable sources, it is carbon neutral and easier to manage than many other fuels. However, industry experience integrating methanol is scarce due to the lack of vessels operating on the fuel.

MethanolPac includes both low- and high-pressure parts of the fuel supply system as well as the related control and safety functions, including the high-pressure methanol fuel pump unit, low-pressure pump module, fuel valve train, bunkering stations and tank instrumentation.

The Wärtsilä 32 methanol engine combines long-established methanol fuel injection technology with control and automation systems. In addition, methanol fuel injection can be retrofitted to any of the Wärtsilä engines in operation. Wärtsilä plans to develop an ammonia-fueled engine concept by 2023 and a hydrogen concept by 2025.

TanQuid, Implico complete digitalization of Duisburg terminal

TanQuid and Implico have completed the digitalization project at the Duisburg tank terminal utilizing Implico's OpenTAS 6.0. The new IT landscape offers automated workflows, expanded service offerings, comprehensive data collection and evaluation, and paperless processes.

The OpenTAS 6.0 automizes all major operations at a facility. It connects with the enterprise resource planning system, gathers data and produces key performance indicators (KPIs). The technology can also be combined with a variety of modules to adapt to the facility's needs.

GAS PROCESSING & LNG

Catherine Watkins, Publisher
+1 (713) 520-4421
Catherine.Watkins@HydrocarbonProcessing.com
www.GasProcessingNews.com

NORTH AMERICA

MIDWEST/CENTRAL U.S.
Josh Mayer
+1 (972) 816-6745
Josh.Mayer@GulfEnergyInfo.com

WESTERN U.S., BRITISH COLOMBIA
Rick Ayer
+1 (949) 366-9089
Rick.Ayer@GulfEnergyInfo.com

NORTHEAST U.S., EASTERN CANADA
Merrie Lynch
+1 (617) 594-4943
Merrie.Lynch@GulfEnergyInfo.com

OUTSIDE NORTH AMERICA

AFRICA
Dele Olaoye
+1 (713) 240-4447
Africa@GulfEnergyInfo.com

BRAZIL
Evan Sponagle
Phone: +55 (21) 2512-2741
Mobile: +55 (21) 99925-3398
Evan.Sponagle@GulfEnergyInfo.com

CHINA
TOPCO
China@GulfEnergyInfo.com

WESTERN EUROPE
Hamilton Pearman
+33 608 310 575
Hamilton.Pearman@GulfEnergyInfo.com

INDIA
Manav Kanwar
+91 (22) 2837 7070
India@GulfEnergyInfo.com

ITALY, GREECE, TURKEY
Filippo Silvera
Office: +39 022846716
Mobile: + 39 3924431741
Filippo.Silvera@GulfEnergyInfo.com

JAPAN
Yoshinori Ikeda
+81 (3) 3661-6138
Japan@GulfEnergyInfo.com

KOREA
YB Jeon
+82 (2) 755-3774
Korea@GulfEnergyInfo.com

RUSSIA, FSU
Lilia Fedotova
+7 (495) 628-10-33
Russia@GulfEnergyInfo.com

UK, SCANDINAVIA, IRELAND, MIDDLE EAST
Brenda Homewood
Phone: +44 (0) 1622 297123
Brenda.Homewood@GulfEnergyInfo.com

CORPORATE, FULL ACCESS SUBSCRIPTION AND CLASSIFIED SALES
J'Nette Davis-Nichols
+1 (713) 520-4426
Jnette.Davis-Nichols@GulfEnergyInfo.com

RE-PRINTS
Lisa Payne with Mossberg & Co.
+1 (219) 561 2036
lpayne@mossbergco.com

Published articles are available for distribution in a PDF format or as professionally printed handouts. Contact Mossberg & Co. for a price quote and details about how you can customize with company logo and contact information.

ADVERTISER INDEX